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Control Technology Study for the New Mexico Potash Industry

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CONTROL TECHNOLOGY STUDY FOR THE NEW MEXICO POTASH INDUSTRY

FINAL

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1201 Elm Street
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DISCLAIMER

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SECTION 1

SUMMARY AND CONCLUSIONS

The particulate concentration of 1.92 gr/dscf determined from this study is significantly lower than previously reported by the potash industry. When sampled with a heated probe and oven to evaporate the water and render the parent particle size of the gas stream (test 0-1), approximately 95 percent of the particulate material is less than 1.3 micrometers aerodynamic diameter. At stack conditions (tests 0-2 and 0-3) the particle size distribution data show that 25 percent of the solids are less than 1 micrometer diameter.

Particulate emission control technology applicable to submerged combustion evaporators used by International Minerals and Chemicals Corporation (IMC) was evaluated in this study. IMC produces various potash products with an annual sales value of $45 million. Nine submerged combustion evaporators are used in the production of sulfate of potash which represents $13.7 million of sales.

The fine particulate emissions from the submerged combustion evaporators can be reduced to 0.03 gr/dscf in an economically and technically feasible manner. Current and future high energy prices indicate that credits for heat recovery will exceed emission control costs. Three control options, Brink mist eliminator, Venturi scrubber and triple effect evaporators, were considered. Required capital investments for the three systems would be $2.5 million, $0.6 million and $4 million, respectively. Annual costs before credits would be about $0.6 million for the Brink system, $0.33 million for the Ventury scrubber and $1.68 million for the triple effect evaporators. Credits for heat recovery would exceed the gross annual costs resulting in net annual credits of $425,000 and $710,000 for the Brink and Venturi systems, respectively. The triple effect evaporator would eliminate the submerged combustion evaporators and result in a net annual credit of $611,000.

Financing of the emission control systems discussed above does not appear to be a problem based on the data available to GCA. The net annual credit that occurs when heat recovery is included represents an attractive investment. In addition, as discussed in Section 1, IMC's sulfate of potash (K_2SO_4) business appears stable, especially since Duval has stopped production.

Future natural gas and electricity costs and availability increase the attractiveness of the triple effect evaporator. The venturi scrubber and Brink mist eliminator options when coupled with submerged combustion evaporators consume relatively large quantities of natural gas and electricity. Costs for these commodities will certainly increase and therefore increase the attractiveness of triple effect evaporators. Triple effect evaporators would use steam produced in boilers that could be designed to fire any fuel. Oil might be used with some problems in the submerged combustion evaporators but coal could not be used.

SECTION 2

INTRODUCTION

Potash was discovered in the Carlsbad area in 1925 and production began in 1935 by the Potash Company of America. International Minerals and Chemical Corporation (IMC) began production in 1940. About 95 percent of U.S. potash consumption is for agricultural purposes. Potash is produced in three chemical forms: muriate of potash (KCl), sulfate of potash-magnesia ($K_2SO_4 \cdot 2\ MgSO_4$) and sulfate of potash (K_2SO_4).[1]

In recent years the U.S. potash industry has received increasing competition from foreign producers as shown in Figure 1.[2,3] Imports currently account for 66 percent of U.S. potash consumption. Canada is the primary source of imports to the U.S. representing 97.2 percent of imports and 64 percent of U.S. consumption.

Sales and price trends in the New Mexico potash industry are presented in Table 1. The data show that production has declined about 20 percent in the past 10 years but prices have more than doubled. After a large price increase from 1974 to 1975 prices have declined slightly. During the 1977 crop year New Mexico accounted for 84 percent of potash sales by U.S. producers.[3]

The primary focus of this study is operations using submerged combustion evaporation. Duval discontinued submerged combustion evaporation during 1971 leaving IMC as the only one used in the New Mexico potash industry. Climax Chemical uses submerged combustion to evaporate salt solutions. IMC currently produces about 250,000 ton/year of KCl, 550,000 ton/year of Sul-Po-Mag (an IMC trademark for $K_2SO_4 \cdot 2\ MgSO_4$) and 150,000 ton/year of K_2SO_4.[4] Current market prices for KCl, $K_2SO_4 \cdot 2\ MgSO_4$, and K_2SO_4 are \$42/ton, \$37.50/ton and \$91/ton, respectively.[5] Total market value of the IMC products if \$45 million/year which the K_2SO_4 product accounting for $\$13.70 \times 10^6$. Production of K_2SO_4 would be most directly affected by any changes to the submerged combustion evaporators as this equipment is used to produce K_2SO_4.

Sales of U.S. produced K_2SO_4 are presented in Table 2. This product is priced over twice as high as muriate of potash (the dominant product on the market). During 1976, 59 percent of the sales of K_2SO_4 were to domestic users. Imports accounted for only 21 percent of domestic consumption with Germany, France, Belgium and Italy accounting for 96 percent of the imports. Potassium sulfate is preferred over muriate of potash for tobacco and crops which will later be distilled.[6]

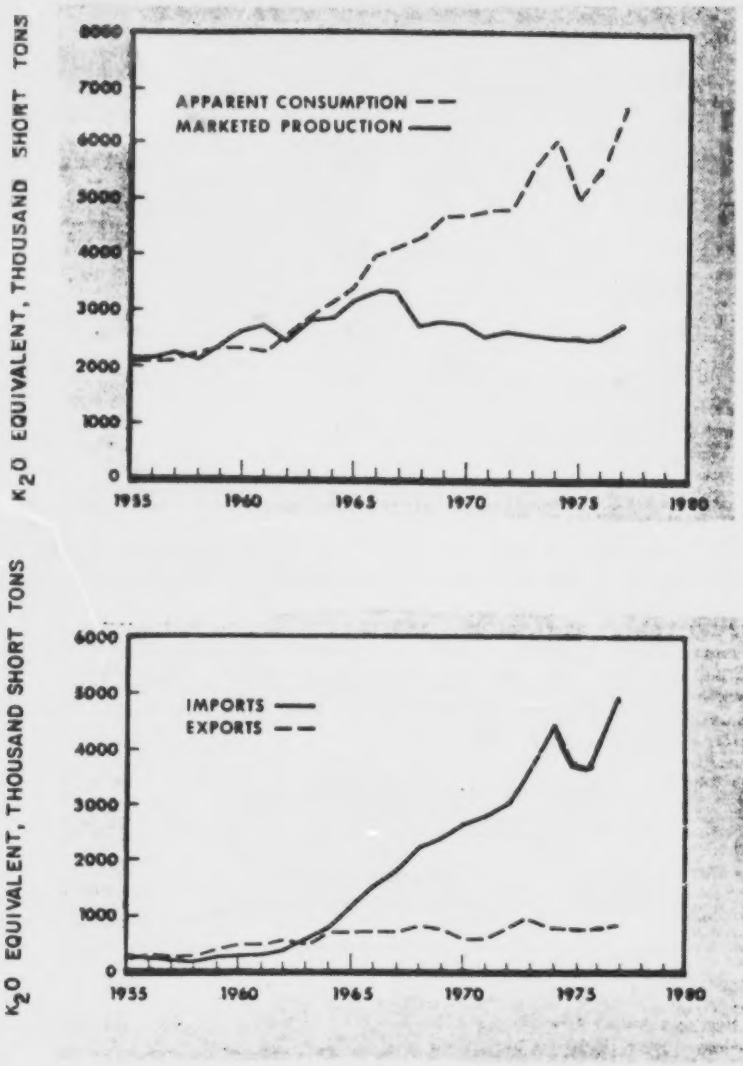

Figure 1. Marketable production, apparent consumption, exports and imports of potassium salts measured in K_2O equivalent (Reference 2, last two points estimates from Reference 3).

TABLE 1. SALES OF NEW MEXICO POTASH*,†

Year	1000 short tons (K_2O equivalent)	Average value ($/tons K_2O equivalent)	Total value ($10^6)
1967	2784	32	89
1968	2511	28	70
1969	2521	26	65
1970	2669	35	92
1971	2285	39	88
1972	2317	39	91
1973	2422	43	103
1974	2061	61	123
1975	1749	86	151
1976	2001	85	169
1977	2187	78	170

*One ton of KCl is equivalent to 0.61 ton K_2O, 1 ton of K_2SO_4 is equivalent to 0.5 ton K_2O and 1 ton $K_2SO_4 \cdot 2\ MgSO_4$ is equivalent to 0.22 ton K_2O.

†Data for years 1967 through 1975 are based on References 2, and 7 through 12 and are for the calendar year. The 1976 and 1977 data are based on Reference 3 and are for the crop year (July 1 through June 20).

TABLE 2. SALES OF U.S. PRODUCED K_2SO_4*

Year	1000 short tons (K_2O equivalent)	Average value ($/tons K O equivalent)	Total value ($10^6)
1972	202	81	16
1973	254	83	21
1974†	237	111	26
1975†	213	179	38
1976†	237	197	47

*References 2,3,10,11, and 12.

†Data are for the crop year.

The market picture for potassium sulfate appears quite different than the outlook for the overall potash industry. There is less competition from imports, especially from Canada. At the present time IMC accounts for about one third of U.S. production. Great Salt Lake Minerals and Chemicals in Ogden, Utah appears to be another major producer and Kerr-McGee in Trona, California also produces sulfate of potash. Prices have been firm and demand in 1978 may exceed 1977.[13,14]

Particulate emissions from submerged combustion evaporators used in the production of sulfate of potash are the subject of this investigation. Submerged combustion evaporators are most applicable where severe corrosion and/or scaling are problems. They have historically been used in the potash industry, the phosphoric acid industry and other miscellaneous applications. Operating submerged combustion evaporators using high efficiency particulate control methods directly applicable to the potash industry could not be located. In the phosphoric acid industry submerged combustion evaporators are being gradually replaced by other systems. Venturi scrubbers have been used in the phosphoric acid industry. A Brinks mist eliminator has reportedly achieved particulate collection efficiencies above 99.4 percent on a submerged combustion evaporator processing $FeCl_2$ in California.[15] Alternative evaporator designs are much more efficient than Ozark evaporators, consuming only 20 to 40 percent as much energy. With current and future energy costs submerged combustion evaporators would probably not be favored at new facilities. Heat recovery equipment however can be applied to submerged combustion evaporators to improve their efficiency.

SECTION 3

PROCESS DESCRIPTION[1,16]

The production of potash at IMC begins with the mining of langbeinite ($K_2SO_4 \cdot 2\ MgSO_4$ and impurities), sylvite (KCl and impurities) and/or mixed langbeinite/sylvite ores. Mining is conducted 700 to 900 feet underground.

The surface operations begin with the crushing of the ores as illustrated in Figure 2. Halite (NaCl) is removed from the crushed langbeinite ore in a countercurrent water leach system (see the upper left-hand portion of Figure 2). After crushing, the mixed ores are treated in the heavy media plant (see upper right-hand corner of Figure 2) where waste halite is removed from the ore and langbeinite and sylvite are separated. The langbeinite from the heavy media plant and the leached langbeinite ore are combined, centrifuged, dryed in a rotary gas-fired dryer and screened to produce the final Sul-Po-Mag products. Sylvite from the heavy media plant is ground in a rod mill and then pressed through a froth flotation system. Finally, the sylvite is centrifuged, dryed and screened to produce the muriate of potash products. The fines from the final Sul-Po-Mag and muriate of potash screening are processed in the chemical plant to produce sulfate of potash. This latter operation is the primary focus of this study.

The production of sulfate of potash and the role of the submerged combustion evaporators are illustrated in Figure 3. Sul-Po-Mag fines are ground in Raymond Mills. The muriate fines are dissolved in water and after thickening they are mixed with the ground Sul-Po-Mag fines in the reaction tanks.

The following reactions occur in the tanks:

$$2\ (K_2SO_4 \cdot 2\ MgSO_4) + 2\ KCl + 12\ H_2O \longrightarrow 3\ (K_2SO_4 \cdot MgSO_4 \cdot 4\ H_2O) + MgCl_2$$

(Langbeinite) (Sylvite) (Water) (Leonite) (Magnesium Chloride)

$$(K_2SO \cdot MgSO_4 \cdot 4\ H_2O) + 2\ KCl \longrightarrow 2\ K_2SO_4 + MgCl_2 + 4\ H_2O$$

Leonite Sylvite Potassium Sulphate Magnesium Chloride Water

Some of the potassium sulfate precipitates and is recovered in centrifuges. The solids from the centrifuges are dryed and screened to produce the final sulfate of potash products.

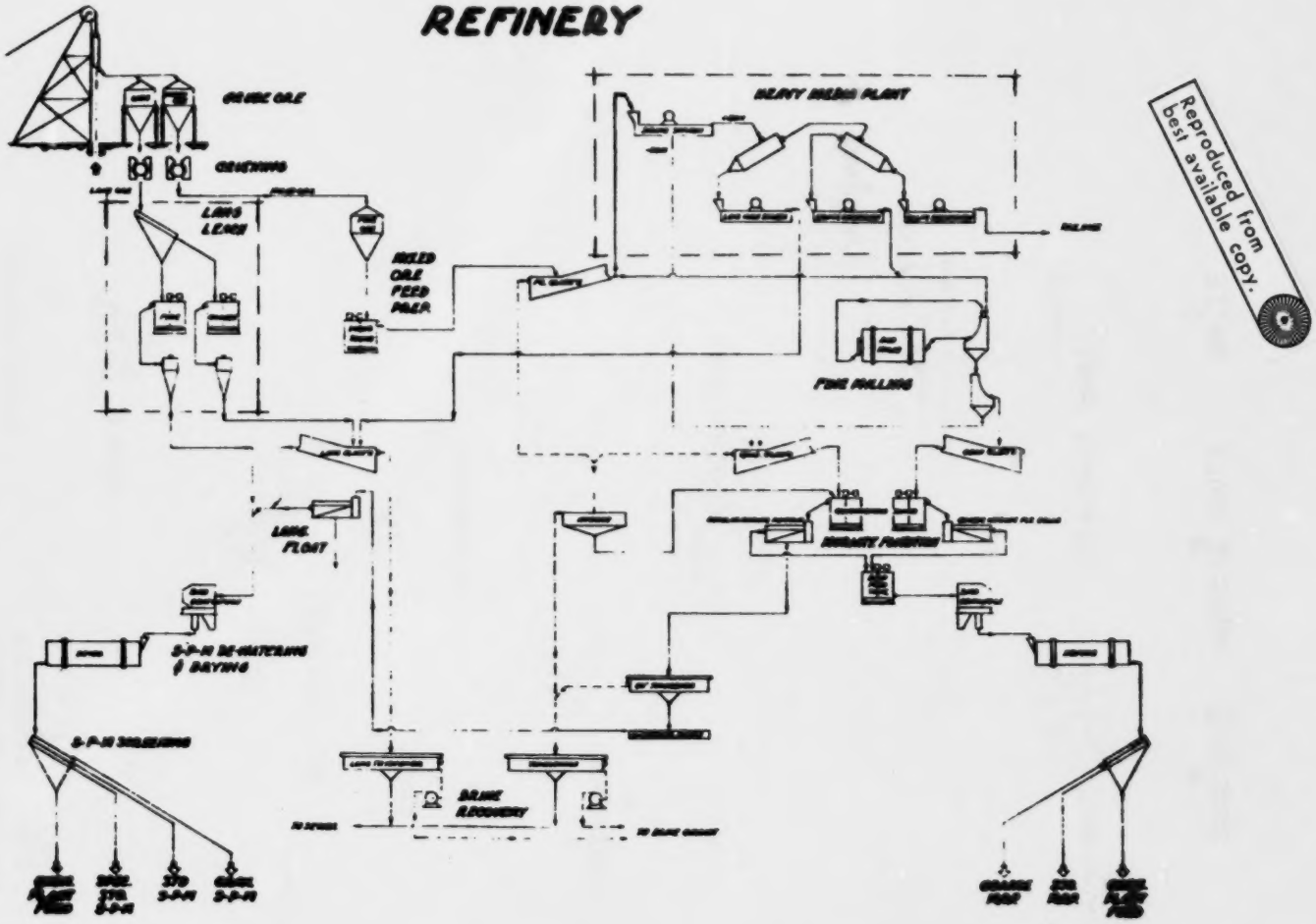

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Figure 2. IMC potash refinery.[1]

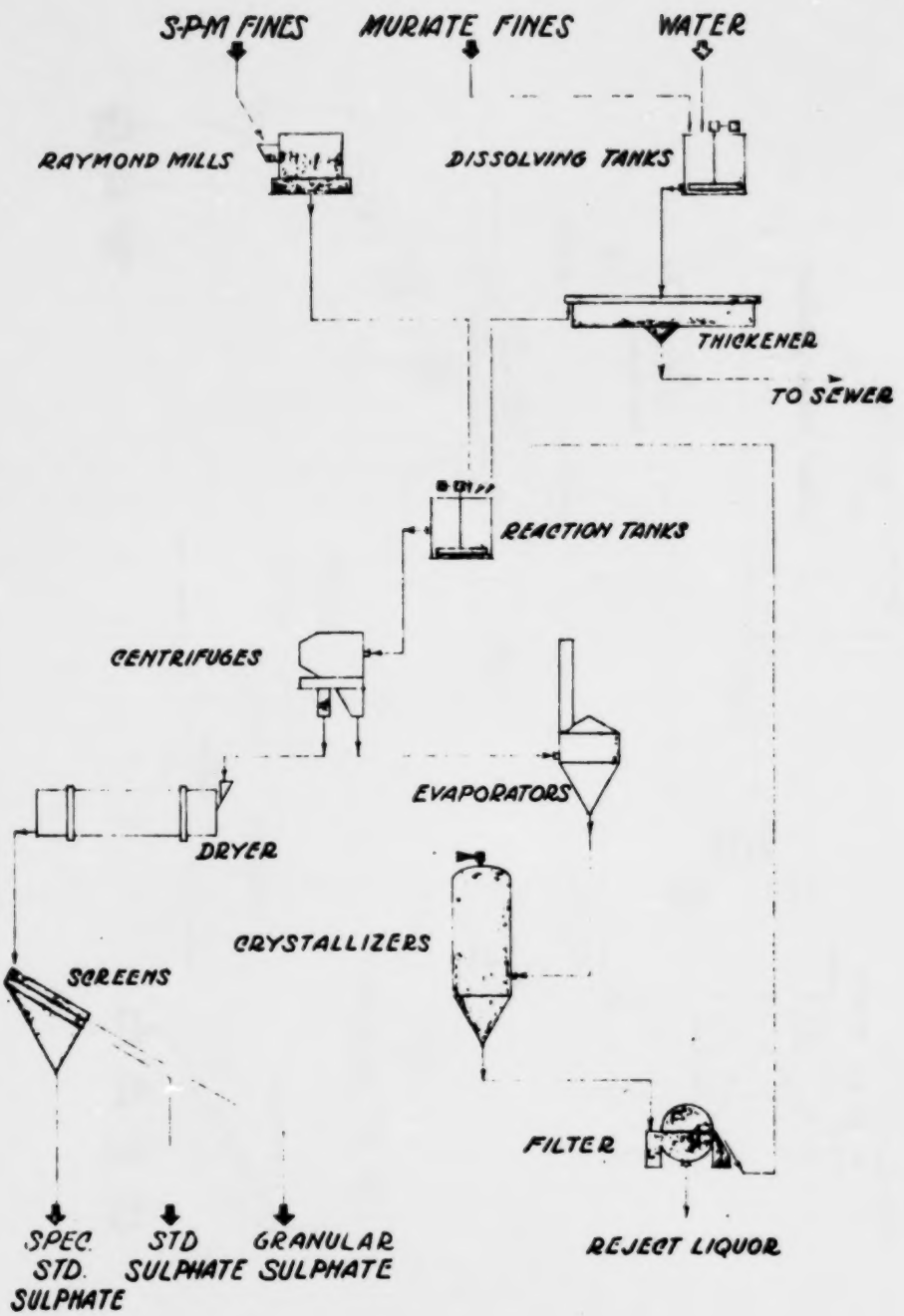

Figure 3. IMC chemical plant for producing K_2SO_4.[1]

Because some potassium remains in the solute from the centrifuges it is economically desirable to recover these potash values. The solute also contains dissolved $MgCl_2$ and smaller quantities of NaCl wastes. In the submerged combustion evaporators the solute is concentrated to precipitate potassium compounds. When NaCl begins to precipitate evaporation is stopped and the solution is cooled. A rotary filter is used to recover potassium values while waste NaCl and $MgCl_2$ leave the system in the reject liquor.

The evaporation step is reportedly needed to effect a reasonable potassium recovery. IMC operates nine submerged combustion evaporators to process about 400 gpm[17] of liquor from the centrifuges. Selected operating parameters are summarized in Table 3. The data show a required evaporation duty of 122,000 lb/hr. In addition, 15 percent of the fuel heating value is lost from incomplete combustion. It should be noted that the above heat loss only applies to the operations as tested by GCA. Also, 12 percent of the input heat is required to raise the liquid temperature. At the current natural gas price of $\$1.56/10^6$ Btu in the Carlsbad area, the nine evaporators use $2.1 million of fuel per year.

TABLE 3. SELECTED OPERATING PARAMETERS - IMC SUBMERGED COMBUSTION EVAPORATORS*

Liquid feed, gal/min	400§
Liquid temperature, °F	131
Liquid specific gravity	1.47
Stack gas temperature, °F	200
Stack gas flow, acfm	100,000
CO content of stack gas, percent vol/vol	0.45†
Natural gas consumed, 10^6 Btu/yr	1,351,000
Heat loss from incomplete combustion, 10^6 Btu/yr	200,000‡
Heat required to raise liquid temperature to 200°F, 10^6 Btu/yr	166,000‡
Heat available for evaporation, 10^6 Btu/yr	975,000‡
Water evaporated, 10^6 lb/yr	975‡
lb/hr	122,000‡
Fuel required per pound of evaporation, Btu/lb	1,372†

*Data collected from IMC unless indicated. Operation at 400 gpm for 8,060 hours per year.

†Measured during tests by GCA.

‡Calculated values.

§Discussion with IMC in June 1977 and emission data in references 4 and 17 for 24-hour and annual periods indicate that operation is constant at 400 gal/min liquid feed and 430 ton/day production. Data showing operations above 400 gal/min and demonstrating the need to provide for a design above 400 gal/min have not been presented to GCA.

SECTION 4

SAMPLING AND ANALYSIS

A series of tests was performed at the IMC plant during the week of December 12 through 16, 1977. The sampling consisted of four in-stack tests performed on the No. 1 Ozark evaporator.* The purposes of the tests were to determine the total particulate emission rate, the volumetric concentration and the particle size distribution. The tests were designed to measure both dry emission and wet particle size distribution.

In order to achieve the desired results, a SASS train was used for the in-stack testing. Although consideration had been given to utilizing in-situ measurement methods rather than extractive techniques, it was decided that the use of a SASS train for particle sizing and total mass sampling would best meet the needs of the program. In-stack impactors offer the advantage of more particle size cuts but are very difficult to use effectively in liquid aerosols. The use of in-stack impactors by IMC has not met with much success because of the very high particulate loading encountered and also possibly due to the hydromagnesite problem. Even if the 3.5 to 4.5 grains per cubic foot loadings reported were dry, presently available impactors could be operated for a sample volume of about 1/2 cubic foot before overloading stages. This sample volume would be reached anywhere from less than 1 minute to perhaps several minutes sampling time and neither the sampling time nor volume could be considered representative. Furthermore, studies by GCA and many other EPA researchers have shown that probe losses with in-stack impactors, even under good sampling conditions, often are in the 25 to 50 percent range. The reasons for these high probe losses are not well understood. Although probe losses are surely encountered in extraction devices, the SASS train utilizes much higher than normal velocities in the probe to minimize probe losses. In this regard, when the probe is operated at stack temperatures, any changes to the sampled aerosol may very well occur further up the stack anyway.

The SASS train is a high volume sampling unit which includes 316 stainless steel cyclones for particle sizing (see Figure 5). The cyclones have nominal cut-points of 10 μm, 3 μm, and 1 μm. In addition, there is a 142 mm filter which collects particles less than 1 μm. The SASS train also includes a gas cooler with an exchange resin for capturing organic emissions and a series of four impingers designed to collect gaseous metals. For the purposes of this program the gas cooler was not required and was omitted. The impingers were

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*A schematic of the stack sampling port is shown in Figure 4.

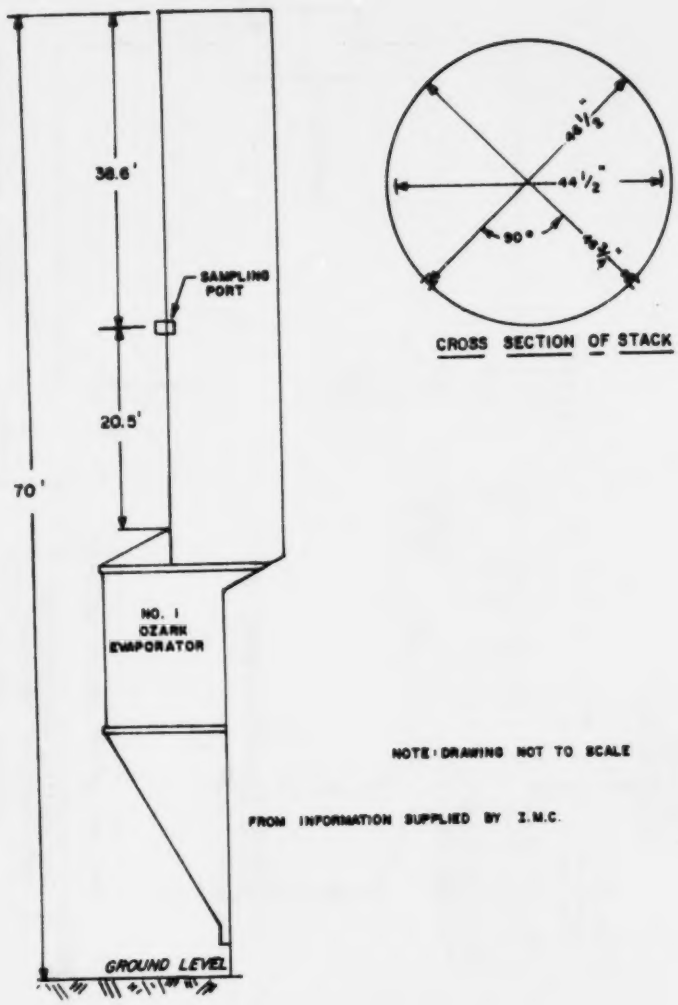

Figure 4. IMC Carlsbad Refinery No. 1 Ozark evaporator stack.

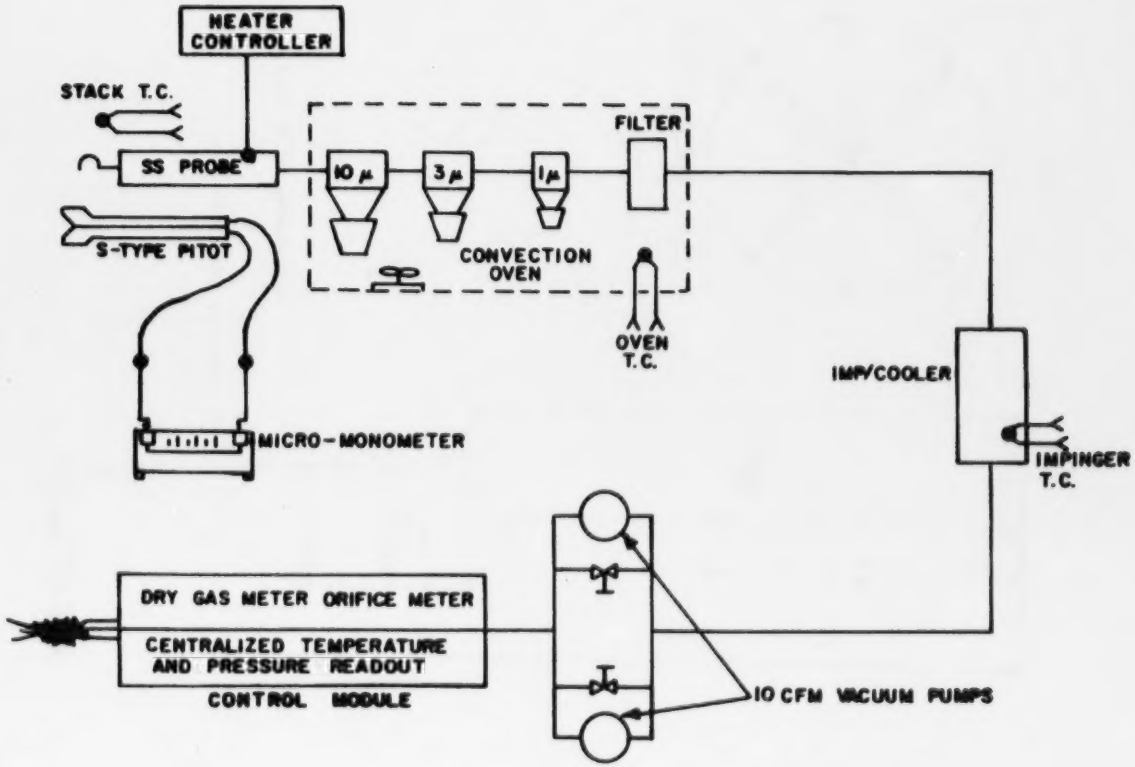

Figure 5. SASS schematic Ozark evaporator IMC.

used initially for the collection of moisture only but later were modified to facilitate testing.

A number of modifications to the sampling train and alterations of the sampling plan were necessary due to problems encountered during the testing. (See Table 4.) The modifications of the sampling train were implemented in order to accomplish the desired goals of the testing program.

Since part of the test was to determine the size distribution of the liquid droplets, two of the tests were conducted with the temperature of the oven at approximately the temperature of the stack gas. The purpose of these tests was to collect the actual droplets in the cyclones in order to determine the size distributions of the liquid particles. The liquid catches were weighed from each cyclone and the percent by weight distribution was calculated for each fraction on the basis of the total moisture content of the test. The results of these tests (0-2, 0-3) are shown in Tables 5 and 6.

In addition to the liquid fraction weighings, the liquid cyclone catches were filtered through a 0.22 μm Millipore filter and then desiccated for dry weights. Each cyclone was also rinsed with acetone after the removal of its liquid and the acetone washes were dried down in a closed hood. The desiccated weights of each of the procedures are included in Tables 7 and 8. These results provide an indication of the solids content of each liquid cyclone catch.

It should be noted that there were a number of problems during these tests which required modifications of the sampling train. The most notable was the clogging of the 142 mm filter. The clogging was presumably due to the hydromagnesite in the stack gas particulate. Thus, while test 0-2 lists two filters being used it was later necessary to eliminate the filters and to install a glass wool trap in the impingers. As a result, the dry data shown in Table 7 is not representative of the actual dry distribution since some portion of the less than 1 μm dry fraction is unaccounted by the modification.

This discrepancy is not true for test 0-3 where a tared glass wool trap was used exclusively for the duration of the test. In this test the net gain of the glass wool plus the filtered impinger water (0.22 μm Millipore) are representative of the less than 1 μm dry fraction. The large gross tare weight of the glass wool trap does limit the precision of the values of the dry weights obtained in this test and the collection efficiency of the glass wool is not known.

An accurate representation of the dry weight particulate size distributions was the object of tests 0-1 and 0-4. In these tests the oven temperature was maintained at 400°F. At the completion of the tests the cyclone catches and filters were desiccated and weighed. Each cyclone and filter housing as well as the probe was brushed and washed with acetone. The washes were dried down in a closed hood and then desiccated and weighed. The results of these two tests are shown in Tables 9 and 10.

Again in these tests the clogging of filters was a major problem. During test 0-1 five filter changes were required to sample for only 16.2 minutes. This was done by using preloaded filter housings. The nozzle wash and brushing were not included in the calculations since it was necessary to leave the probe

TABLE 4. OZARK EVAPORATORS, IMC, DESCRIPTION AND DATA OF TESTS

Test No.	Date	T_{oven} (^{o}F)	Total time (min)	V_{mstd} (dscf)	V_{water} (scf)	Description of train	Percent I	$Q_{cyclone}$ (acfpm)
0-1	12/13/77	400	16.2	24.632	44.85	Three cyclones plus filter, changed five times	111.5	7.98
0-2	12/14/77	200	3.5	7.433	10.388	Three cyclones plus filters (2), replaced by glass wool trap in impinger No. 3	133.1	7.33
0-3	12/15/77	111	5.0	9.623	14.008	Three cyclones, no filter, glass wool in impinger No. 3	130.7	5.74
0-4	12/15/77	405	34.0	49.290	74.15	Three cyclones, no filter, glass wool in impinger No. 3	101.4	6.67

TABLE 5. OZARK EVAPORATORS IMC TEST 0-2, WET FRACTION

Particulate sample	Wet catch (g)	Percent of total*
1 μm	65.564	4.1
3 μm	18.6887	8.5
10 μm	8.9621	29.8

*Percent wet total as $\frac{\text{g wet catch}}{406.8 \text{ g water total}}$

TABLE 6. OZARK EVAPORATORS IMC TEST 0-3, WET FRACTION

Particulate sample	Wet catch (g)	Percent of of wet total*
1 μm	11.6646	3.9
3 μm	16.3125	5.5
10 μm	55.7451	18.8

*Percent wet total as $\frac{\text{g wet catch}}{297.1 \text{ g water total}}$

TABLE 7. OZARK EVAPORATORS IMC - TEST 0-2

Particulate sample	Dry catch (g)	Acetone wash (g)	Total (g)	Percent of total
Filters (2)	0.4049	0.0118	0.4167	82.9
1 μm	0.0027	0.0052	0.0079	1.6
3 μm	0.0044	0.0007	0.0051	1.0
10 μm	0.0033	0.0146	0.0179	3.6
Probe	0	0.0549		
Total catch			0.5025	

TABLE 8. OZARK EVAPORATORS IMC - TEST 0-3

Particulate sample	Dry catch (g)	Acetone washes (g)	Total (g)	Percent of total	
1 μm	0.0341	0.0038	0.0379	3.2	
3 μm	0.0009	0.0032	0.0041	0.3	
10 μm	0.0008	0.0112	0.012	1.0	
Glass wool	1.0		1.0	85.2	< 1 μm size
Impingers filter	0.0008		0.0008	0.1	
Probe			0.1183	10.1	
Total catch			1.1731		

TABLE 9. OZARK EVAPORATORS IMC - TEST 0-1

Particulate sample	Catch (g)	Acetone rinse (g)	Total (g)	Percent of total
Filters (5)	2.96814	0.0152	2.98334	94.7
1 μm	.0102	0.0438	0.0540	1.7
3 μm	0.0	0.0162	0.0162	0.5
10 μm	0.0	0.0119	0.0119	0.4
Probe	0.6	0.0865	0.0865	2.7
Total catch			3.15194	

TABLE 10. OZARK EVAPORATORS IMC - TEST 0-4

Particulate sample	Dry catch (g)	Acetone rinse (g)	Total (g)	Total (dry)	
1 μm	0.2003	0.0192	0.2155	9.7	
3 μm	0.0308	0.0433	0.0741	3.3	
10 μm	0.4062	0.0130	0.4192	18.6	
Probe		0.6658	0.6658	29.9	
Glass wool	0.8		0.8	39.9	< 1 μm size
Impingers filter	0.0913		0.0913	2.3	
Total catch			2.2259		

in the stack during the filter changes. The probe was left in the stack during filter changes due to the problem of the physical replacement of the probe-oven unit into the stack. This problem was the result of limited space on the stack platform, bulk nature of SASS assembly, high CO and positive stack pressure. The nozzle was heavily caked with the stack particulate. These washes were recovered but not included in any calculations since the probe was in the flue gas for a much longer time than was actually sampled. It was decided that the recovered amount would grossly alter the grain loading while the exclusion of the nozzle wash should be insignificant compared to the large filter catches. The results shown in Table 9 indicate that the major fraction of the dry particulate was less than 1 μm.

On the basis of this information, test 0-4 was designed to provide further information on the other size fraction distributions. The filter was eliminated and a tared glass wool trap was placed in impinger No. 3 of the train. With this design it was assumed that the volume of gas samples could be increased by collecting the less than 1 μm fraction on the glass wool trip thus avoiding the rapid plugging experienced with the filters. The results are shown in Table 10. The impinger waters were also filter with a 0.22 μm Millipore filter and these data are included as part of the less than 1 μm fraction. The results of this test are questionable due to the low total particulate grain loading value which may indicate that the glass wool trap was inefficient over the longer sampling period. Also, the high percentage of dry catch in the probe is of unknown size distribution.

The grain loading calculations showed some variability but the average value for tests 0-1 and 0-3 of 1.92 gr/dscf (Table 11) is considered to be an accurate value for total flue gas particulate loading. The two low values for tests 0-2 and 0-4 are assumed to be the results of loss of the less than 1 μm size fraction due to the train modifications. These values are all lower than were expected but no explanation for any discrepancy is available although process operating parameters may be responsible. Size distribution data is assumed to be an accurate representation of the actual particulate emissions in spite of the numerous problems and modifications. The D_{50} calculations for each test are shown in Table 12 and in Figure 6 a graphic representation of this data is presented with the percent by weight correlation.

Also presented in Table 11 are the results of Orsat analysis of the integrated bag samples taken each day during the testing. The variation between each day is unexplained and plant data was unavailable to confirm the readings. Values presented were assumed representative of the gas composition during each test and these values were used for all corresponding calculations. In tests 0-3 and 0-4 the single bag sample results were used for both tests since they were both completed within the same day.

The results of the testing indicate a grain loading of 1.92 gr/dscf with an average of 85 percent of the size fraction as less than 1 μm particles.* The

*The particle emission size distribution information for the less than 1 μm diameter is available only as an extrapolation of the figures shown in Figure 6 but comparison of this extrapolation to the date obtained by Monsanto Enviro-Chem shows good agreement.

TABLE 11. OZARK EVAPORATORS, IMC. – TEST PARAMETERS

Test #	T_{stk} average (^{o}F)	% moisture	% CO_2	% O_2	% CO	Gas flow (dscfm)	Grain loading (gr/dscf)	Gram/min (actual)
0-1	205	64.5	8.0	4.6	0.9	2938.4	1.971	375.3
0-2	207	58.3	9.1	4.0	0.8	3487.3	0.903	204.1
0-3	182	59.3	9.6	3.6	0.6	3218.3	1.878	391.6
0-4	214	61.5				3124.4	0.696	140.9

TABLE 12. OZARK TESTS – CYCLONE DIAMETER CUTS D_{50}

Test #	$Q_{cyclone}$ (acfpm)	μ*	Diameter (μm)		
			Small cyclone	Medium cyclone	Large cyclone
0-1	7,98	0,022	1,23	2.96	8.41
0-2	7.31	0.017	1,13	2.72	7.72
0-3	5.74	0.015	1.20	2.89	8.19
0-4	6.67	0.022	1.35	3.25	9.19

*μ is calculated gas viscosity (centipoise).

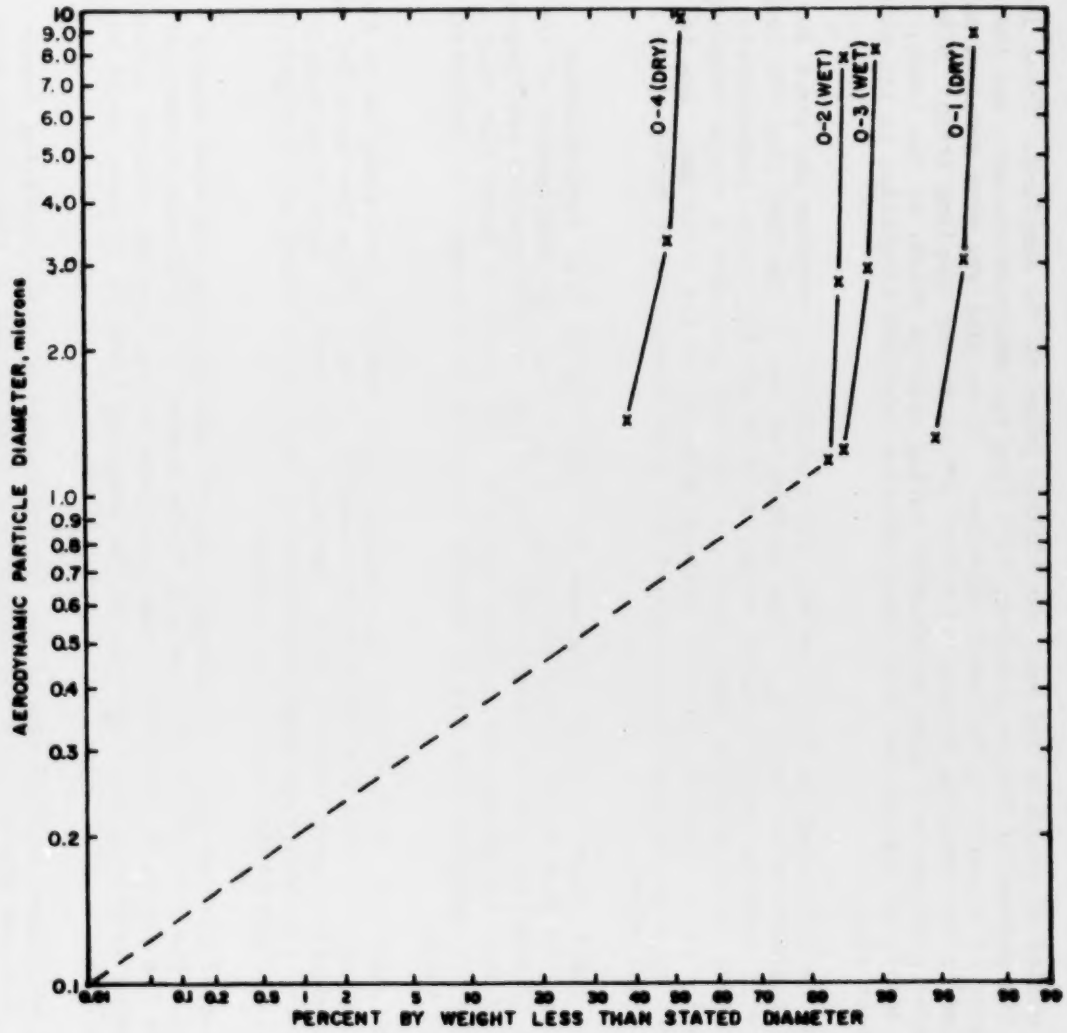

Figure 6. Ozark evaporator stack No. 1 particulate emission size distributions using SASS train.

average stack gas CO content was determined to be 0.45 percent (V/V) and the average moisture content of the gas was 57 percent. This data along with design data collected from IMC has been incorporated into the control technology evaluation section.

SAMPLING PROBLEMS DISCUSSION

A number of problems had been anticipated prior to the sampling. These included the problem of filter clogging, high flue gas moisture content, and the occurrence of droplets in the stack emissions. These problems were remedied as described in the sampling methods by modifications in the sampling train. Other unexpected problems were also encountered during testing which, at the least, caused delays in testing and as a final solution required alteration in the proposed test plan.

The major problem in this area was the inability to traverse the stack for a total representative sampling. This problem was due to the fact that the large nozzle size required for isokinetic sampling would not fit in the perpendicular ports. As a result a third larger diameter port was used with a single sampling point selected as representative of average velocity of the stack gas. Due to the short duration of the tests it would have been impossible to traverse the stack in any event.

The short duration of the test was the result of the high loading rates which resulted from rapid filter plugging. As a result, the maintenance of the desired ΔH sampling rate was impossible. The effect of this problem was frequent filter changes and anisokinetic sampling rates. It should be noted that test 0-4 for which the sampling time was 34 minutes resulted in an acceptable isokinetic sampling rate.

The expected effect of the hyperanisokinetic sampling rate would be to bias the percentage of small particles in the sampled volume. Since the tests 0-2 and 0-3 were designed such that the small particles measurements were of secondary importance, the resulting hyperisokinetics of these tests would not be significant in the interpretations of the results.

Another problem which continually hindered the sampling was power supply failures. The SASS train requires a multiple number of independent electrical circuits for complete operation. Numerous times during testing, various circuits would break requiring shutting down of the sampling train until power could be restored.

An additional problem was in the interpretation of plant operating conditions. This was particularly noticeable in the monitoring of the orifice pressure drop readings which, while they were continuous, it was not known if these were representative of normal operating conditions. Time chart recordings were available but normal operating parameters were not defined.

The test results presented in this report are considered representative of the stack flue gas at the conditions of operating during the test. All problems encountered were remedied to the satisfaction of the testing program in order to provide representative data.

SECTION 5

CONTROL TECHNOLOGY

The submerged combustion evaporators used by IMC produce particulate emissions in relatively large amounts. Tests show that, by weight, most of the particles (solid or liquid), are less than 1 micrometer diameter (see Section 3). Numerous control options could be considered. In this study three options as listed below were considered:

- Brink* Mist Eliminator
- Venturi Scrubber
- Alternative Evaporators

Within each of the major control options listed above it would be possible to evaluate many specific heat recovery methods, equipment design variations, and emission goals. In this study an emission goal of 0.03 gr/dscf was established. Venturi scrubbers, Brink mist eliminators and alternative evaporator designs appear to be technically capable of achieving 0.03 gr/dscf.

Monsanto, Duval and IMC have tested a pilot scale Brink mist eliminator at Duval.[16,17,19] The following discussion is based primarily on the above tests and personal communications with the companies involved. A schematic of a Brink mist eliminator is presented in Figure 7. The Brink mist eliminator consists of patented fiber beds. Higher packing densities increase collection efficiencies but also increase the pressure drop across the system. Monsanto literature indicates that efficiencies over 99 percent are possible for all particle sizes. The system was developed by Monsanto primarily for their own sulfuric and phosphoric acid plants. Over 1800 Brink systems have been installed worldwide.[20]

Monsanto Enviro-Chem, Duval and IMC have tested a pilot scale Brink mist eliminator at Duval's plant. With a sulfuric acid wash, operation was satisfactory and emissions were reduced to 0.03 gr/dscf. Heat recovery by a packed tower was used in the pilot tests.

Cooling the total submerged combustion evaporator flue gases from 190°F to 100°F would decrease their heat content by about 1,000,000 × 10^6 Btu/yr. If

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*Trademark of Monsanto Enviro-Chem Systems, Inc., Box 14547, St. Louis, Missouri 63178.

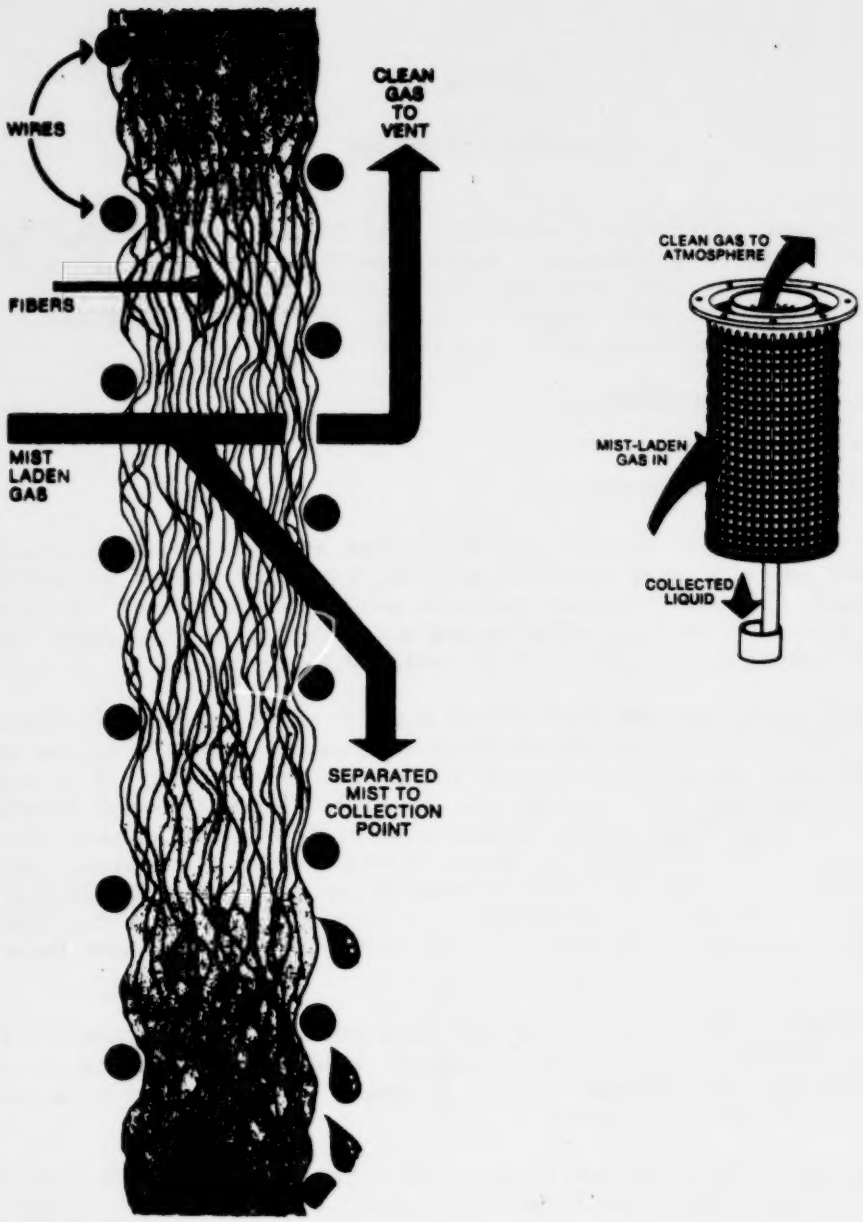

Figure 7. Brink mist eliminator.[20]

heat can be effectively recovered and utilized the potential cost savings in natural gas would be $1,560,000 per year. Some of the heat might be used to preheat the feed to the Ozark evaporators so that more material could be processed in a single evaporator. Also heat could be used in a vacuum evaporator (if one were purchased) thus reducing the load on the Ozark evaporator. Capital equipment and operating costs would need to be compared to the energy savings to determine the best system. It should be noted that energy savings should not necessarily be treated as a result of installing emission controls. For instance, IMC could install a packed tower heat recovery system achieving only minimal control of particulates but recovering large amounts of energy.

Capital and operating costs for a Brinks mist eliminator system are presented in Table 13. The costs would include a packed tower to aid in heat recovery and particulate control. No credit is included for heat recovery. The capital cost estimate is based on data provided by Duval[19] indicating a cost of $500,000 for a system 1/5 the size required for IMC and communications with IMC[16,17] and Monsanto.[21] It should be noted that IMC considers the $2,500,000 cost as appropriate only for a very simple system. Monsanto estimated $400,000 for the Duval system but cautioned that a different system might be needed for IMC. The bases for the operating costs are presented in the footnotes to Table 13. An annual cost of $615,000 was calculated for this system.

A Venturi scrubber was the second alternative evaluated. This system would also include a packed tower for heat recovery and any coincidental particulate removal or reduction of potential particle growth. However, data are inadequate to predict the effect of a packed tower on the submicron particulates and its effect was not included in the analysis. One approach used to predict the performance of a Venturi scrubber was a recently published mathematical model.[22] This mathematical model has been developed over several years by Air Pollution Technology Inc. EPA's major contractor for wet scrubber research. While there is some uncertainty associated with the use of this mathematical model, (or any model) the APT model is generally considered to be the best available. Because the model neglects diffusion, it may, however, underestimate the collection of particles below 0.5 microns. An inlet concentration of 1.92 gr/dscf was used based on measurements by GCA discussed previously. At the selected liquid-to-gas ratio of 0.002 ft^3/ft^3 and the GCA measured size distribution, a pressure drop of over 100 in. W. C. would be required to reduce outlet concentration to 0.03 gr/dscf. The size distribution determined by Monsanto[34] apparently at Climax Chemical would require a pressure drop of 68 in. W.C.

A Venturi scrubber is currently operating at Climax and on a submerged combustion evaporator processing NaCl brine. Test data are not yet available but reported visual observations indicate that opacity has been sharply reduced from about 100 percent to 20 percent.[19] In addition, the manufacturer, Hydronics Engineering, has provided a guarantee to achieve 0.03 gr/dscf.[27] This system included a packed tower prescrubber and operates at a pressure drop of 40 in. W.C.

TABLE 13. COSTS FOR BRINK MIST ELIMINATOR, 10^3

Capital investment*	2,500
Annual operating cost	
Labor (including maintenance)†	63
Administrative and general overhead‡	38
Materials§	63
Electricity#	28
Sulfuric acid**	38
Local taxes and insurance††	68
Net operating cost	298
Annualized Cost	
Annualized capital cost‡‡	274
Federal income tax§§	43
Net operating cost	298
Total annual cost	615

*Capital costs based on personal communication with Don Felton, IMC, Mike Scroggin, Duval, and Dave Duros, Monsanto. The cost represents a relatively simple system.

†Labor is primarily for maintenance and assumed by GCA to equal 2.5 percent of the capital investment. This assumption agrees with standard engineering practices.[23]

‡Administrative and general overhead are assumed by GCA to be equal to 60 percent of labor. This is the high end of the reported range of 40 to 60 percent.[23]

§Materials for maintenance are assumed by GCA to equal the labor cost.[23,24]

#Electricity cost is based on the prevailing industrial rates in the Carlsbad area of 2.6¢/kWh.[26] Also, power is assumed to be primarily for the system pressure drop of 16 in. W.C. Electricity consumption is calculated from the system pressure drop.[25]

**Based on the projected consumption of 1500 ton/yr and a delivered cost of $25/ton for recovered acid.[17]

††2.7 percent of capital investment.[24]

‡‡Based on a 20-year useful life and a discount cash flow method with an interest rate of 9 percent. This cost accounts for interest and capital recovery.

§§1.73 percent of the capital investment.[24]

For the purpose of this analysis we have assumed that a Venturi scrubber pressure drop of 70 in. W.C. would enable IMC to achieve 0.03 gr/dscf. Several factors were considered in estimating the installed capital cost. First, the unit at Climax Chemical which is designed for 30,000 acfm cost $100,000 installed.[28] Parts of this scrubber are rubber-lined and the Venturi throat is stainless steel. Second, capital costs in the literature are generally in the $2 to $8/acfm range. For this study a cost of $6/acfm was used. It should be noted that the scrubber capital cost has only a minor impact on annual costs because of the high energy consumption as illustrated in Table 14. Total annual cost for the Venturi scrubber system is about $330,000.

The third alternative considered was to replace the existing submerged combustion evaporators with more efficient triple effect evaporators. A schematic of a triple effect evaporator is presented in Figure 8. This type of system and many variations are used to achieve high heat efficiencies. Water vapor generated in one stage is used to evaporate more water in the next stage. One pound of steam can be used to evaporate 3 or more pounds of water.[29] Evaporator fuel consumption could be reduced from present levels by about 70 percent. Particulate emissions would be negligible.

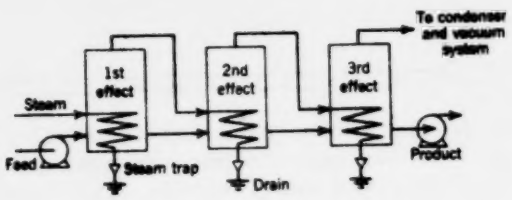

Figure 8. Schematic of a forward feed triple effect evaporation system.[30]

The capital costs for triple effect evaporator system were based on a brief discussion with Mr. J.D. Dockendroff of HPD, Inc.* Mr. Dockendroff indicated that HPD recently submitted a prepared quotation for a quadruple effect evaporator for a NaCl application. The above system was designed for a heat economy of 3.4 (1 pound of steam will evaporate 3.4 pounds of water) and would be constructed of titanium and monel. About 130,000 lb/hr of water would be evaporated. The evaporators would cost about $2 million and the total project cost would be $3-1/2 million.[31]

At IMC the required evaporator duty would be 122,000 lb/hr (see Table 12). Also, 41,000 lb/hr gas-fired boiler would be needed to generate steam. Therefore, the estimated total capital cost is $4 million. Annual costs would be $1,680,000 as shown in Table 15.

---

*HPD Inc., Glen Ellyn, Illinois.

TABLE 14. COSTS FOR VENTURI SCRUBBERS, 10^3

Capital investment (installed cost)	600*
Annual operating cost	
Labor including maintenance	21†
Administrative and general overhead, 60 percent of labor	12‡
Materials	21†
Electricity	184#
Local taxes and insurance	16**
Net operating cost	254
Annualized cost	
Annualized capital cost	65††
Federal income tax	11‡‡
Net operating cost	254
Total annualized cost	330

*Based on typical costs in the technical literature and the complexity of controlling nine sources. Also, the cost of the scrubber at Climax ($3.3/acfm was considered). The selected capital cost is $6/acfm.

†Labor is primarily for maintenance and is assumed by GCA to equal 2.5 percent of the capital cost. This assumption agrees with standard engineering practices.[23]

‡Administrative and general overhead are assumed by GCA to equal 60 percent of the labor cost. This is the high end of the reported range of 40 to 60 percent.[23]

§Materials are for maintenance and are assumed by GCA to equal the labor cost.[23,24] It is further assumed that a waste liquid stream is used for scrubbing.

#Based on the gas flow of 100,000 acfm (51,000 acfm at fan), 8,060 hours of operation/year and on a typical Carlsbad area electricity cost of 2.6¢/kWh.[19] Electricity consumption is calculated from the system pressure drop.[25]

**2.7 percent of capital investment.[24]

††Based on a 20-year useful life and a discount cash flow method with an interest rate of 9 percent.

‡‡1.73 percent of the capital investment.[24]

TABLE 15. COSTS FOR A TRIPLE EFFECT EVAPORATOR TO REPLACE OZARK EVAPORATORS, 10^3

Capital investment (installed cost)	4,000*
Annual operating costs	
Labor (including maintenance)	140†
Administration and general overhead	84‡
Materials	140§
Natural gas	602#
Local taxes and insurance	108**
Miscellaneous operating costs	100††
Net operating costs	1,184
Annualized costs	
Annualized capital cost	438‡‡
Federal income tax	69§§
Net operating cost	1,184
Total annual cost	1,681

*Based on a discussion with HPD, Inc.; see text.[25]

†Labor is primarily for maintenance and is assumed by GCA to be equal to 3-1/2 percent of the capital investment. This assumption agrees with standard engineering practices.[23] The cost is also higher than the 2-1/2 percent value used for the Brink Mist Eliminator.

‡Administrative and general overhead costs are assumed by GCA to equal 60 percent of the labor costs. This is the high end of the reported range of 40 to 60 percent.[23]

§Materials are for maintenance and are assumed to equal 3-1/2 percent of the capital investment. Based on references 23 and 24 and GCA's evaluation of system complexity.

#The cost for natural gas is based on boiler efficiency of 85 percent, an evaporator duty of 122,000 lb/hr with a heat economy of 3 and 8,060 hours of operation. The typical price for natural gas in the Carlsbad area is $1.56/10^6$ Btu.[33]

**2.7 percent of the capital investment.[24]

††Included to account for water and miscellaneous electrical costs.

‡‡Based on a 20-year useful life and a discount cash flow method with an interest rate of 9 percent.

§§1.73 percent of capital investment.[24]

Financial data that could be used to determine the impact of the costs presented from the Brink mist eliminator, Venturi scrubber and triple effect evaporator are not available. Relative costs that are available, however, are discussed below. First, the annual costs for the Brink system is about $615,000; the annual cost for the Venturi scrubber is $330,000; and the annual cost for the triple effect evaporator is $1,680,000. With the first two systems the potential for recovery of heat which is about $1,560,000 can be considered. The triple effect evaporator would eliminate the cost of operating the Ozark evaporator.

Our estimate of Ozark operating costs are presented in Table 16. Actual costs may be higher but this conservative lower estimate is adequate for the present discussion. The Ozark evaporators were installed in the early 1950's when natural gas only cost $0.20/$10^6$ Btu compared to present costs of $1.56/$10^6$ Btu.[32,33] Our cost estimates indicate that IMC could reduce annual evaporation costs by $600,000 by installing triple effect evaporators and eliminate particulate emissions at the same time. Even if the required capital investment for triple effect evaporators was $6 million (50 percent higher than our estimated $4 million) the annual cost would be $2.2 million or approximately equivalent to the present cost.

The Brink mist eliminator and Venturi scrubber costs can also be compared to the Ozark evaporator costs. Annual costs for the Brink mist eliminator and Venturi scrubber (not including credits for heat recovery) are 27 and 14 percent of the Ozark operating cost. It should be noted, however, that the Ozark operating cost is only part of the cost of producing sulfate of potash.

Potential control costs can also be compared to sales reveneus to obtain an indication of the impact. As previously noted, the total market value of IMC's products is $45 million. Therefore, the annual costs of the Brink system are equivalent to 1.4 percent of sales and the cost of the Venturi scrubber are equivalent to 0.7 percent of sales. The evaporation step at IMC is used to produce sulfate of potash with a current annual market value of $13.7 million. Control costs for the Brink and Venturi systems are respectively equivalent to 4.4 percent and 2.4 percent of the sulfate of potash market value. Also, it should be noted that IMC spent $35 million for expansion of production in 1974.[12]

Some pertinent cost data are summarized in Tables 17 and 18. There is a net reduction in costs for each system after the credits for reduced operating costs are considered. The potential energy recovery with the Brink and Venturi systems is $1,500,000. Net annual costs for these two systems assuming that two-thirds of the heat is recovered are -$425,000 and -$710,000 for the Brink and Venturi systems, respectively. As mentioned previously, the net annual costs for the triple effect evaporator are -$611,000.

TABLE 16. OZARK EVAPORATOR COSTS, 10^3

Capital investment	0
Annual operating costs	
Labor	70*
Administrative and general overhead	42*
Materials	7*
Natural gas	2,110†
Local taxes and insurance	0
Net operating cost	2,292
Annualized costs	
Annualized capital cost	0
Federal income tax	0
Net operating cost	0
Total annualized cost	2,292

*Assumed equal to 50 percent of the costs for the triple effect evaporation system.

†Based on the fuel use reported in Reference 4 and the typical natural gas cost in the Carlsbad area of $1.56/10^6$ Btu.[33]

TABLE 17. SUMMARY OF CONTROL EQUIPMENT COSTS, 10^3

	Brink system	Venturi scrubber	Triple effect evaporator
Capital investment	2,500	600	4,000
Annual cost (before credits)	615	330	1,681
Potential credits	1,560*	1,560*	2,292†
Net costs	-425‡	-710‡	-611

*Potential credits for heat recovery. It is unlikely that this total heat value would be recovered.

†Credits for elimination of Ozark evaporator operating costs.

‡ Assuming that two-thirds of the heat is recovered.

TABLE 18. COMPARISON OF CONTROL COSTS TO IMC SALES, 10^3

Annual cost before credits	
Brink system	615
Venturi scrubber	330
Triple effect evaporator	1,681
Annual cost after credits	
Brink system	-425*
Venturi scrubber	-710*
Triple effect evaporator	-611
Ozark operating cost	2,292
Sulfate of potash sales	13,700
Total potash sales	45,000

*Assuming that two-thirds of the heat is recovered.

APPENDIX A

RUN DATA SHEETS

PRELIMINARY DATA

OZARK EVAPORATOR NO. 1

IMC

12/12/77

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PLANT IMC

DATE 12/02/77

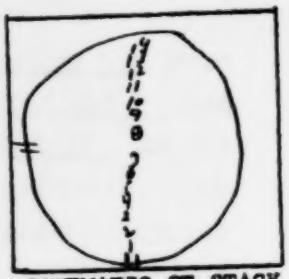
RUN NO. Prelim Traverse

STACK DIAMETER, in. 46 3/4

BAROMETRIC PRESSURE, in.Hg.

STATIC PRESSURE IN STACK(P_g), in. Hg.

OPERATORS WP, RG



SCHEMATIC OF STACK CROSS SECTION

South Port

Traverse point number	Velocity head, in. H_2O	$\sqrt{\Delta_p}$	Stack Temperature(T_s), °F
1	.025	.1581	185
2	.03	.1732	190
3	.04	.200	195
4	.045	.2121	197
5	.045	.2121	198
6	.05	.2236	199
7	.055	.2345	200
8	.060	.2550	200
9	.070	.2646	200
10	.060	.2550	200
11	.070	.2646	200
12	.070	.2646	201
13	.060	.2550	200
14	~~AVERAGE~~ .055	.2345	199
	Average	.229	197.

STAT.C = .12

PLANT IMC

DATE 12/2/77

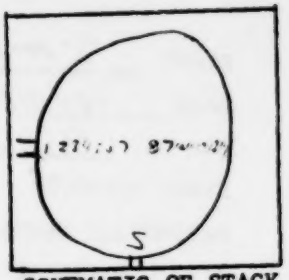
RUN NO. Prelim. Traverse

STACK DIAMETER, in. 48

BAROMETRIC PRESSURE, in.Hg.

STATIC PRESSURE IN STACK(P_g), in. Hg.

OPERATORS RMB, WP, RS



SCHEMATIC OF STACK CROSS SECTION

West Port

Traverse point number	Velocity head, in. H_2O	$\sqrt{\Delta_p}$	Stack Temperature(T_S), °F
1	.035	.187	165
2	.04	.200	185
3	.045	.212	187
4	.05	.236	192
5	.06	.249	195
6	.06	.249	196
7	.065	.255	197
8	.07	.265	197
9	.06	.249	197
10	.06	.249	197
11	.065	.255	198
12	.065	.255	198
13	.055	.235	198
14	~~AVERAGE~~ .04	.200	198
	Average	.2523	194

ORSAT

DATE

Work - 12/12/79

PRELIMS

INLET 0.7L.T

TIME _____

GAS	ORSAT READINGS		Avg	% GAS
CO_2	8.9	8.7	8.8	8.8
O_2	12.9	13.0	13.0	4.2
CO	13.5	13.7	13.6	0.6
				N_2 = 100 - 87.4

OUTLET _____

TIME _____

GAS	ORSAT READINGS	%GAS
CO_2		
O_2		
CO		

(samples from Int Bag Sample)

3' 10"

$m_s = .44\,(CO_2) + .32\,(O_2) + .28\,(N_2 + CO)$

$m_s = .44\,(8.8) + .32\,(4.2) + .28\,(87.0$

TEST NO. 0-1

OZARK EVAPORATOR NO. 1

IMC

12/13/77

SASS ANALYTICAL DATA

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PLANT IMC / [illegible]

DATE 12/13/77

SAMPLING LOCATION —

SOURCE I.D. —

RUN NUMBER 0-1

RECOVERY BY

RECORDED

COMMENTS: each component rinsed separately with acetone after recovery wt to be added to total. for particulate

FILTER # 0-15 0.[illegible]7461
0-14 0.[illegible]
0-13 0.[illegible]
0-12 0.[illegible]
0-11 0.37279
T.W. 2.96814

CYCLONES USED:

10 x None 3 x 2 Watts 1 x 0.0102 gm
31.0614
31.0716

rinses

VOLUME

	INITIAL	FINAL
FIRST IMPINGER	300 ml	>500 ml
SECOND IMPINGER	300 ml	470 ml
THIRD IMPINGER	0 ml	0 ml
TOTAL	600 ml	1220 ml

added to silica gel because of back flow during power failure

GAIN 620 ml

CONDENSATE ~~TRAP~~ GAIN ml

SILICA GEL WEIGHT

INITIAL	FINAL
750.0 g	1081.3 g
g	g
g	g
TOTAL g	TOTAL g

GAIN 331.3 g

TOTAL 951.3 ml

ADDITIONAL COMMENTS:

188.455

GCA/TECHNOLOGY DIVISION

SASS FIELD DATA

PLANT _____
DATE _____
SAMPLING LOCATION _____
SOURCE I.D. _____
RUN NUMBER I
OPERATOR _____
AMBIENT TEMPERATURE _____
BAROMETRIC PRESSURE _____
STATIC PRESSURE, (P_s) _____
FILTER NUMBER(S) _____

PROBE LENGTH AND TYPE _____
NOZZLE I.D. _____
ASSUMED MOISTURE, % _____
XAD-2 MODULE NUMBER _____
METER BOX NUMBER _____ OVEN NUMBER _____
METER $\Delta H_@$ _____
C FACTOR _____
PROBE HEATER SETTING _____
OVEN SETTING _____
REFERENCE ΔP _____

SCHEMATIC OF TRAVERSE POINT LAYOUT
READ AND RECORD ALL DATA EVERY ___ MINUTES

TRAVERSE POINT NUMBER	SAMPLING TIME, MIN	CLOCK TIME (24 HR CLOCK)	GAS METER READING (V_m), ft^3	VELOCITY HEAD (Δp_s), IN. H_2O	ORIFICE PRESSURE DIFFERENTIAL (ΔH), IN H_2O DESIRED	ORIFICE PRESSURE DIFFERENTIAL (ΔH), IN H_2O ACTUAL	STACK TEMPERATURE (T_s), °F	DRY GAS METER TEMPERATURE INLET ($T_{M_{IN}}$) °F	DRY GAS METER TEMPERATURE OUTLET ($T_{M_{OUT}}$) °F	PUMP VACUUM IN Hg	OVEN TEMPERATURE °F	IMPINGER TEMPERATURE °F	SORBENT TRAP TEMPERATURE °F	PROBE TEMPERATURE °F
1	0	1315	203.420	.095	.55	.55	200	84	82	10	400	82	—	405
Stop	2	5		212,						24				
2		14.06	212.835	.035	55		211	84	81	11	403	76		200
Stop #		10	14.10	220,847						23				
7		1456	220,730	.035	.55		206	84	81	14	401	72		396
Stop #		12.3		225.030						21				
7		15.2	225,217	.050	.55	.55	203	76	78	10	401	77		396
Stop #		18.8		229.955						23				
7		15:47	230.163	.045	.55	.53	203	7	7	10	402	74		432
Stop		16.2		232.985										
TOTAL		●		●	●	●	●	●	●	●	●	●	●	●
AVERAGE	●	●	●	●	●					●	●	●	●	●

COMMENTS Stopped @ 18.10 Couldn't maintain ΔH

232.985
203.420
29.565

Imp = .093

.145 = Trial train

38

ORSAT DATE

12/3/77 JMC Run 1 SDSS.

INLET _____ TIME _____

GAS	ORSAT READINGS		% GAS
Co_2	8.0	7.9	8.0
O_2	12.6	12.5	4.6
CO	13.4	13.5	0.9

OUTLET _____ TIME _____

GAS	ORSAT READINGS	%GAS
CO_2		
O_2		
CO		

GCA/TECHNOLOGY DIVISION

BURLINGTON ROAD, BEDFORD, MASSACHUSETTS 01730 / PHONE: 617-275-9000

SHEET____ OF____

BY____

DATE____

CH'K. BY____

DATE CH'K.____

B'K. CH'K. BY____

JOB NO.____

PROJECT (2)

SUBJECT____

12/13/77

Inc

1345 - while changing SASS Filters

gas 5.5 air 8.7

1440 gas 5.8 air 8.7 23 amps @ 2300 volts

1545 gas 5.8 air 8.7 23 - 24 amps

TEST NO. 0-2

OZARK EVAPORATOR NO. 1

IMC

12/14/77

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SASS ANALYTICAL DATA

PLANT TVA C

DATE 2/14/77

SAMPLING LOCATION

SOURCE I.D.

RUN NUMBER 2

RECOVERY BY

RECORDED

COMMENTS:

FILTER # 0-1G 4049
0-3G 0
Glass wool plug

CYCLONES USED:

10μ 65.5649	3μ 18.6877	1μ 8.9621	
24.1612	31.8919	38.7992	Empty
96.7261	92.5772	40.7013	+ contents
			Not
			+ rinse
			.7

VOLUME

	INITIAL	FINAL	
FIRST IMPINGER	ml	450 ml	
SECOND IMPINGER	ml	270 ml	
THIRD IMPINGER	ml	0 ml	
TOTAL	600 ml	720 ml	GAIN 120 ml

~~CONDENSATE~~ - TRAP GAIN ml

SILICA GEL

	INITIAL WEIGHT	FINAL	
	g	757.1 g	1021.1
	g	g	1014.0
	g	g	7.1
TOTAL	g	TOTAL g	GAIN 7.1 g
			TOTAL 127.1 ml

ADDITIONAL COMMENTS:
Acetone in 1 ℓ (brown color)
water in 3-ℓ (dk particulate, small amount)
Acetone in 1 ℓ (brown color, dk particulate pieces)
condensate in 0-3G + filter housing

rinsed w/ acetone after transferring catches to tared 6 oz Nalgene containers

GLASS WOOL = 25.60g
(OVER)

GCA/TECHNOLOGY DIVISION

SASS FIELD DATA

PLANT IMC
DATE 12-14-77
SAMPLING LOCATION
SOURCE I.D.
RUN NUMBER 0-2
OPERATOR
AMBIENT TEMPERATURE
BAROMETRIC PRESSURE 25.91
STATIC PRESSURE, (P_s) -.15
FILTER NUMBER(S) 19, 20

PROBE LENGTH AND TYPE 5'
NOZZLE I.D. .40
ASSUMED MOISTURE, % 6.4
XAD-2 MODULE NUMBER —
METER BOX NUMBER — OVEN NUMBER —
METER ΔH —
C FACTOR —
PROBE HEATER SETTING —
OVEN SETTING —
REFERENCE ΔP

SCHEMATIC OF TRAVERSE POINT LAYOUT
READ AND RECORD ALL DATA EVERY ____ MINUTES

TRAVERSE POINT NUMBER	CLOCK TIME (24 HR CLOCK) / SAMPLING TIME, MIN		GAS METER READING (V_M), FT^3	VELOCITY HEAD (ΔP_s) IN. H_2O	ORIFICE PRESSURE DIFFERENTIAL (ΔH), IN H_2O		STACK TEMPERATURE (T_s), °F	DRY GAS METER TEMPERATURE		PUMP VACUUM IN Hg	OVEN TEMPERATURE °F	IMPINGER TEMPERATURE °F	SORBENT TRAP TEMPERATURE °F	PROBE TEMPERATURE °F
					DESIRED	ACTUAL		INLET ($T_{M_{IN}}$) °F	OUTLET ($T_{M_{OUT}}$) °F					
7	0	10:15	235.602		.9									
	.8		236.955											
7		10:58	236.750		.9		194			10	200			200
	1.5		238.900	.055						23				
7	4	12:06	239.031	.055	.9		215			8.5				★
	.7		240.898							23	—			—
7		13:23	240.942		.9	.9	212	51	81	10	91	76	—	204
	2.0		244.819	.055	.9	.9				21				
TOTAL	●	●	●	●	●	●	●	●	●	●	●	●	●	●
AVERAGE	●	●	●	●	●	●	●	●	●	●	●	●	●	●

COMMENTS 10:15 shut down load power to one pump VAC reach 6" Hg.
10:59 shut down Teflon clogged.
★ Probe temp readout incorrect. 1. Probe temp repaired

Trans = .047
A = Hi-V Clock Reset Button
43
Leak Rate Imp = .021

ORSAT DATE

Run 2 12/14/77

INLET TIME

GAS	ORSAT READINGS	% GAS
CO_2	7.1 7.1	7.1
O_2	13.0 13.2	13.0
CO	13.9	0.0

OUTLET TIME

GAS	ORSAT READINGS	%GAS
CO_2		
O_2		
CO		

JOB NO. _____ SHEET ____ OF ____
PROJECT _____ BY _____
SUBJECT _____ DATE _____
_____ CH'K. BY _____
DATE CH'K. _____
B'K. CH'K. BY _____

stainless steel cylinder
with parafilm on both ends

= 30.30 grams

Gross

= 71.60 grams

41.30

Stainless steel cylinder
w/ parafilm on both ends
After test = 71.65 grams

TEST NO. O-3

OZARK EVAPORATOR NO. 1

IMC

12/15/77

SASS ANALYTICAL DATA

PLANT TMC

DATE 12/15/77

SAMPLING LOCATION

SOURCE I.D.

RUN NUMBER 0-3

RECOVERY BY

RECORDED

COMMENTS:

FILTER # None

CYCLONES USED: 48.6735 41.7200 Final Tot. Net

10 @ 30.8358 3 @ 32.3610 1 @ 30.0554

VOLUME

	INITIAL	FINAL
FIRST IMPINGER	[illegible] ml	435 ml
SECOND IMPINGER	360 ml	358 ml
THIRD IMPINGER	ml	ml
TOTAL	600 ml	793 ml

190.40 g [illegible] / 191.4 wt / dry

GAIN 193 ml

CONDENSATE - TRAP GAIN 8.4 ml

SILICA GEL WEIGHT

INITIAL	FINAL
750.0 g	762.0 g
g	g
g	g
TOTAL g	TOTAL g

1028.6
1016.6
12.0

GAIN 12.0 g

TOTAL ml

ADDITIONAL COMMENTS:

.35 ΔH hot
.6 ΔH cold

GCA/TECHNOLOGY DIVISION

BASS FIELD DATA

PLANT Imc
DATE 12-12-77
SAMPLING LOCATION
SOURCE I.D.
RUN NUMBER #9
OPERATOR
AMBIENT TEMPERATURE
BAROMETRIC PRESSURE 28.32
STATIC PRESSURE, (P_s)
FILTER NUMBER(s)

PROBE LENGTH AND TYPE 6'
NOZZLE I.D. .40"
ASSUMED MOISTURE, %
XAD-2 MODULE NUMBER
METER BOX NUMBER OVEN NUMBER
METER ΔH
C FACTOR
PROBE HEATER SETTING 200°F
OVEN SETTING 0
REFERENCE ΔP

SCHEMATIC OF TRAVERSE POINT LAYOUT

READ AND RECORD ALL DATA EVERY ____ MINUTES

TRAVERSE POINT NUMBER	CLOCK TIME (24 HR CLOCK) / SAMPLING TIME, MIN		GAS METER READING (V_m), ft^3	VELOCITY HEAD (Δp_s), IN. H_2O	ORIFICE PRESSURE DIFFERENTIAL (ΔH), IN H_2O		STACK TEMPERATURE (T_s), °F	DRY GAS METER TEMPERATURE		PUMP VACUUM IN Hg	OVEN TEMPERATURE °F	IMPINGER TEMPERATURE °F	SORBENT TRAP TEMPERATURE °F	PROBE TEMPERATURE °F
					DESIRED	ACTUAL		INLET ($T_{m_{IN}}$), °F	OUTLET ($T_{m_{OUT}}$), °F					
1	0	0930	245.071	.05	.6		175	65	59	8	—		—	175
Stop	.6		246.032							23				
7			246.057											
Stop	.9		246.623	.04	.6									
7			249.090	.05	.6	.6	181	81	69	5	105	68		150
	4		256.150				188	82	71		117	70		178
Stop	5		258.168	.05	.6	.6								
TOTAL		●		●	●	●	●	●	●	●	●	●	●	●
AVERAGE	●	●	●	●	●					●	●	●	●	●

COMMENTS # could not hold ΔH of .6

Final Leak Rate: 258.412

249.071
4.
253.
2.
58
49
9

258.998
258.760
.021 CFM @ 20" Hg

Leak Rate = .05 CFM / 10 sec @ 20" Hg

48

GCA/TECHNOLOGY DIVISION

BURLINGTON ROAD, BEDFORD, MASSACHUSETTS 01730 / PHONE: 617-275-9000

SHEET ____ OF ____

BY ____

DATE 12-15-77

JOB NO. ____

PROJECT ____

SUBJECT Zinc

CH'K. BY ____

DATE CH'K. ____

B'K. CH'K. BY ____

0815 190.90 grams of glass wool is placed in #3 impinger. #1 & 2 imp. have 300 ml of H_2O.
Train will be run w/out a filter in the oven. Oven Temp will be off.
Probe temp 200° F

0820 started test Brought ΔH to .6 But could not Maintain shut down after 6 ½ of 5 min.

0825 started test again without changing glass wool Run time 3rd of 1 min

0840 Disconnected imp. #3 & 4 from Train, Run pressure drop test across Silica gel - glass wool impingers, Could not maintain ΔH of .6

0845 Ran Pressure Drop test across Silica gel Results 9.0 ΔH

0850 Removed some glass wool from imp pressure drop ok.

1000 Ran test held ΔH @ .6 for 4 min 1st test
Shut down test because filters will collect H_2O.
Don't want H_2O in ogelbees to Blow down

#1	#2	#3
246.022	246.673	258.364
245.071	246.057	249.090
.951	.616	9.274

9.274
.616
.961

Total Vol = 10.851

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TEST NO. 0-4

OZARK EVAPORATOR NO. 1

IMC

12/15/77

SASS ANALYTICAL DATA

PLANT Inc
DATE 12/13/77
SAMPLING LOCATION
SOURCE I.D.
RUN NUMBER 0-1
RECOVERY BY
RECORDED

COMMENTS:

FILTER # None

CYCLONES USED: 31.2187 32.0811 Final Tare Wt
10 x 32.4497 3 x 31.1879 1 x 31.8808

VOLUME

	INITIAL	FINAL
FIRST IMPINGER	300 ml	4148 ml
SECOND IMPINGER	300 ml	793 ml
THIRD IMPINGER	— ml	54 ml
TOTAL	600 ml	1995 ml

100.0g dry glass wool 222.4 (wet) after

GAIN 1395 ml

CONDENSATE - TRAP GAIN ml

SILICA GEL WEIGHT

INITIAL	FINAL
7500. g	8053 g
g	g
g	g
TOTAL g	TOTAL g

GAIN 553 g

TOTAL ml

ADDITIONAL COMMENTS:

GCA/TECHNOLOGY DIVISION

SASS FIELD DATA

PLANT IMC
DATE 12-14-77
SAMPLING LOCATION
SOURCE I.D.
RUN NUMBER 0-4
OPERATOR
AMBIENT TEMPERATURE
BAROMETRIC PRESSURE 28.64
STATIC PRESSURE, (P_s)
FILTER NUMBER(s)

PROBE LENGTH AND TYPE 6'
NOZZLE I.D. .4D"
ASSUMED MOISTURE, %
XAD-2 MODULE NUMBER
METER BOX NUMBER OVEN NUMBER
METER ΔH
C FACTOR
PROBE HEATER SETTING 400
OVEN SETTING 400
REFERENCE ΔP

317.625
239.547
77.078

SCHEMATIC OF TRAVERSE POINT LAYOUT
READ AND RECORD ALL DATA EVERY ____ MINUTES

TRAVERSE POINT NUMBER	SAMPLING TIME, MIN	CLOCK TIME (24 HR CLOCK)	GAS METER READING (V_M), FT^3	VELOCITY HEAD ($\sqrt{\Delta P_s}$)^2, IN. H_2O	ORIFICE PRESSURE DIFFERENTIAL (ΔH), IN H_2O DESIRED	ORIFICE PRESSURE DIFFERENTIAL (ΔH), IN H_2O ACTUAL	STACK TEMPERATURE (T_S), °F	DRY GAS METER TEMPERATURE INLET ($T_{M_{IN}}$)°F	DRY GAS METER TEMPERATURE OUTLET ($T_{M_{OUT}}$)°F	PUMP VACUUM IN Hg	OVEN TEMPERATURE °F	IMPINGER TEMPERATURE °F	SORBENT TRAP TEMPERATURE °F	PROBE TEMPERATURE °F
9	0	1245	259.647	.05	.35	.35	219	99	102	5	406	92	—	408
	2		263.145	.055	.35	.35	222	96	99	5.5	415	88	—	399
	4		267.0	.055	.35	.35	221	93	98		416	81	—	398
	6		270.850	.055	.35	.35	221	92	96	6.0	411	80	—	258
	8		274.380	.055	.35	.35	220	91	95	6.0	398	79	—	399
	10		276.760	.055	.35	.35	218	90	94	6.0	398	80	—	402
	12		280.160	.05	.35	.35	216	90	94	6.0	397	86	—	398
	14		283.575	.055	.35	.35	218	90	93	6.0	403	90	—	403
	16		286.850	.055	.35	.35	216	90	93	6.5	410	90	—	397
	18		290.400	.055	.35	.35	215	90	92	6.0	410	100	—	396
Stop —	A	2145	294.745											254
	20.5		298.920	.055	.35	.35	210	90	91	6.5	404	62	—	406
	22.6		297.6	.05	.35	.35	210	90	91	6.5	412	62		400
	24		300.945	.05	.35	.35	210	91	90	7.0	408	64		398
	26		304.395	.055	.35	.35	209	91	90	8.5	397	66		396
	28		307.810	.055	.35	.35	210	92	90	8.5	395	66		392
	30		311.110	.05	.35	.35	210	93	90	7	400	64		406
	32		314.475	.05	.35	.35	209	94	90	14	407	63		400
	34		317.625	.05			214			6.2				
											405 avg			
TOTAL		●		●	●	●	●	●	●	●	●	●	●	●
AVERAGE	●	●	●	●	●			91.7	93.3	●	●	●	●	●

COMMENTS A - Lost Power to probe STop test

VAC. Build up

Leak Rate = .015 CFM @ 20 in Hg / 6.39L

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ORSAT

DATE

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0 0-4 2 hr

INLET _____ TIME _____

GAS	ORSAT READINGS	% GAS	
CO_2	7.6 9.2 7.8 9.4	9.6	9.6
O_2	11.6 13.0 11.7 13.2	13.3	3.6
CO	13.2 13.4 11.7 13.8	13.8	0.6

OUTLET _____ TIME _____

GAS	ORSAT READINGS	%GAS
CO_2		
O_2		
CO		

M_d = 32.0(.136) − .114(9.6) + .28(86.8)

M_d = 4.35 + 4.22 + 24.30 = 29.67

SASS TRAIN CALIBRATIONS

EPA SASS UNIT
"J" FACTOR CALIBRATION

1-11-78

ΔH (in H_2O)	V_1 (cm)	V_2 (cm)	V_2-V_1 (cm)	Θ (min)	t_1 (°F)	t_2 (°F)	P_M (in Hg)	Q_{MD} (CFM)	J_i
1	300.00	314.70	14.70	5	92	81	30.28	2.74	1.04
2	316.60	336.85	20.25	5	99	84	30.28	4.05	1.01
4	365.70	393.25	27.55	5	109	92	30.28	5.51	0.961
6	421.00	464.18	33.18	5	118	101	30.28	6.64	0.928

$\bar{J} = 0.99$

Where $$J_i = \frac{Q_{MD}}{A_{0i}}\left[\frac{(T_{M_{AVG}}+460)}{P_M M_M} \times R \times \frac{\Delta H}{13.6} \times 2g_c\right]^{-1/2}$$

$$Q_{MD} = \frac{V_2 - V_1}{\Theta} = (CFM)$$

$$A_{0i} = \frac{IN^2}{144}$$

P_M = in Hg

$M_M = 29$

$R = 1545$

$g_c = 1:5,812$

EPA SASS UNIT

ΔH_2

ORIFICE DIAM. = 0.358

1-4-78

ΔH ("H$_2$O)	V_1 (CF)	V_2 (CF)	θ MIN	t_1 (°F)	t_2 (°F)	$V_2 - V_1$ (CF)	Q_M (CFM)	K_M	P_M (in Hg)	ΔH
.25	287.40	297.70	5	87	79	10.5	2.08	3.75	30.28	0.51
1.0	300.00	314.70	5	92	81	14.70	2.91	3.70	30.28	1.02
2.0	316.60	336.85	5	99	84	20.25	3.99	3.58	30.28	2.0
3.0	338.70	363.01	5	106	88	24.11	4.74	3.46	30.28	3.0
4.0	365.70	393.25	5	105	92	27.55	5.43	3.42	30.28	4.1
5.0	394.60	425.00	5	114	96	30.40	5.98	3.36	30.28	5.2
6.0	421.00	464.18	5	118	101	33.18	6.54	3.34	30.28	6.3

$\bar{K_M} = 3.52$

Where: $$Q_M = \frac{V_2 - V_1}{\theta}\left[\frac{t_2 + 460}{\left(\frac{t_2 + t_1}{2}\right) + 460}\right]$$

$$K_M = Q_M \left(\frac{P_M M_M}{T_M \Delta H}\right)^{1/2}$$

where: $P_M = P_{stm}$

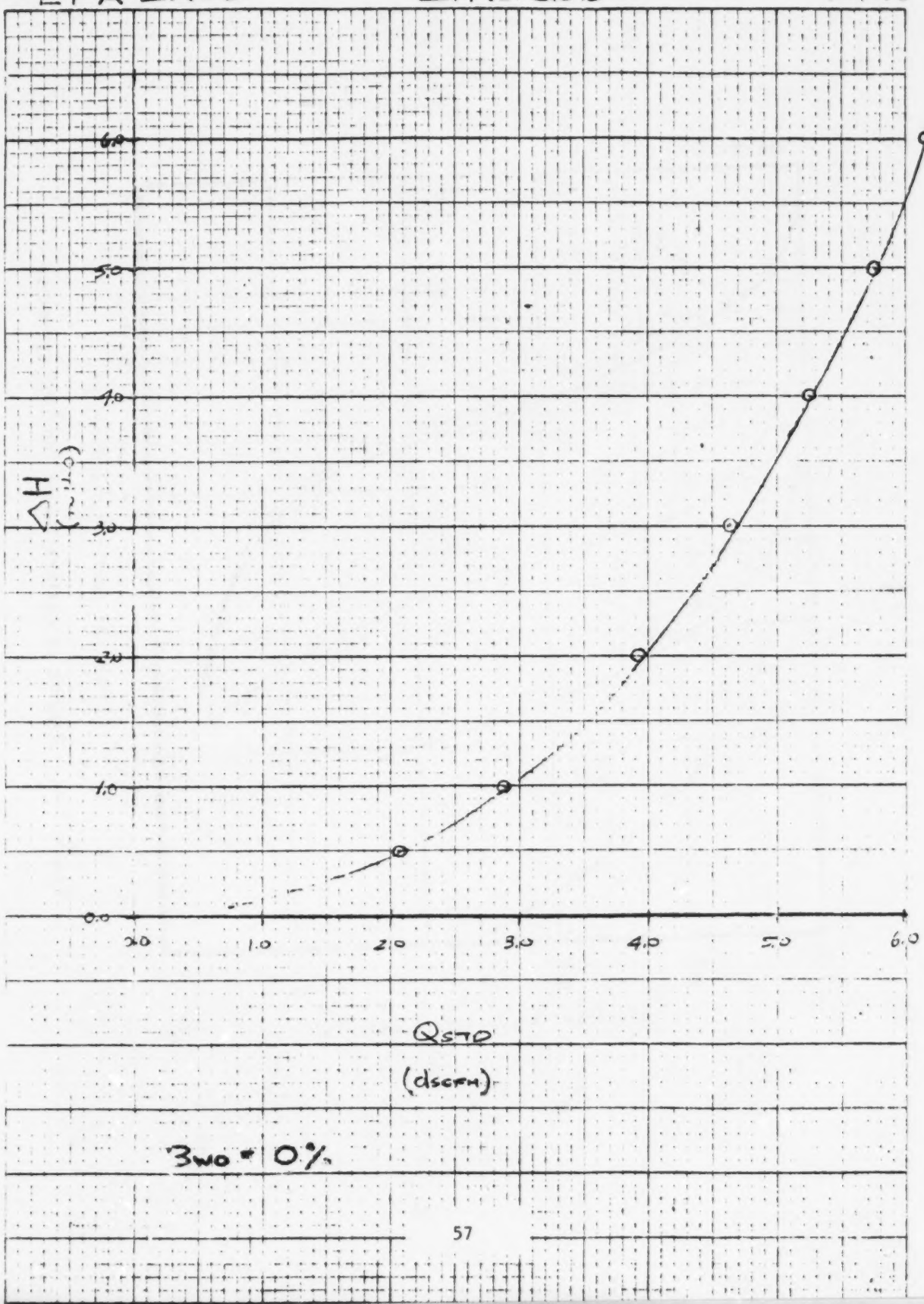
$M_M = M_{(air)} = 2$

$T_M = t_2 + 4$

$$\Delta H_2 = \frac{(Q_M)^2 (29.92)(M_M)}{528} \times \frac{1}{K_M^2}$$

$\therefore Q_M < 0.75$ CFM

EPA SASS ΔH vs Qstd 1-4-78



Bwo = 0%

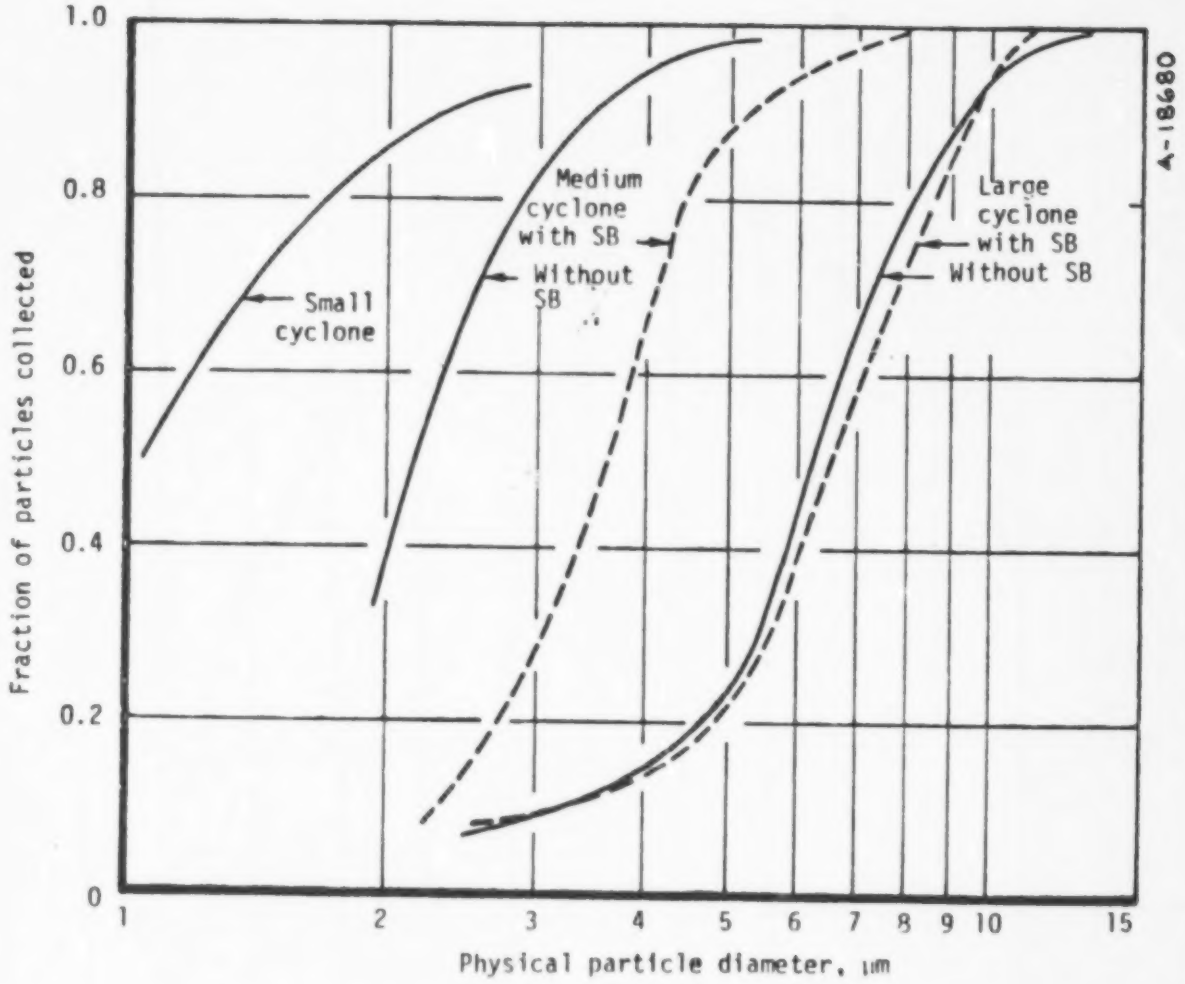


Figure 7-14. Results of calibration of EPA SASS cyclones.

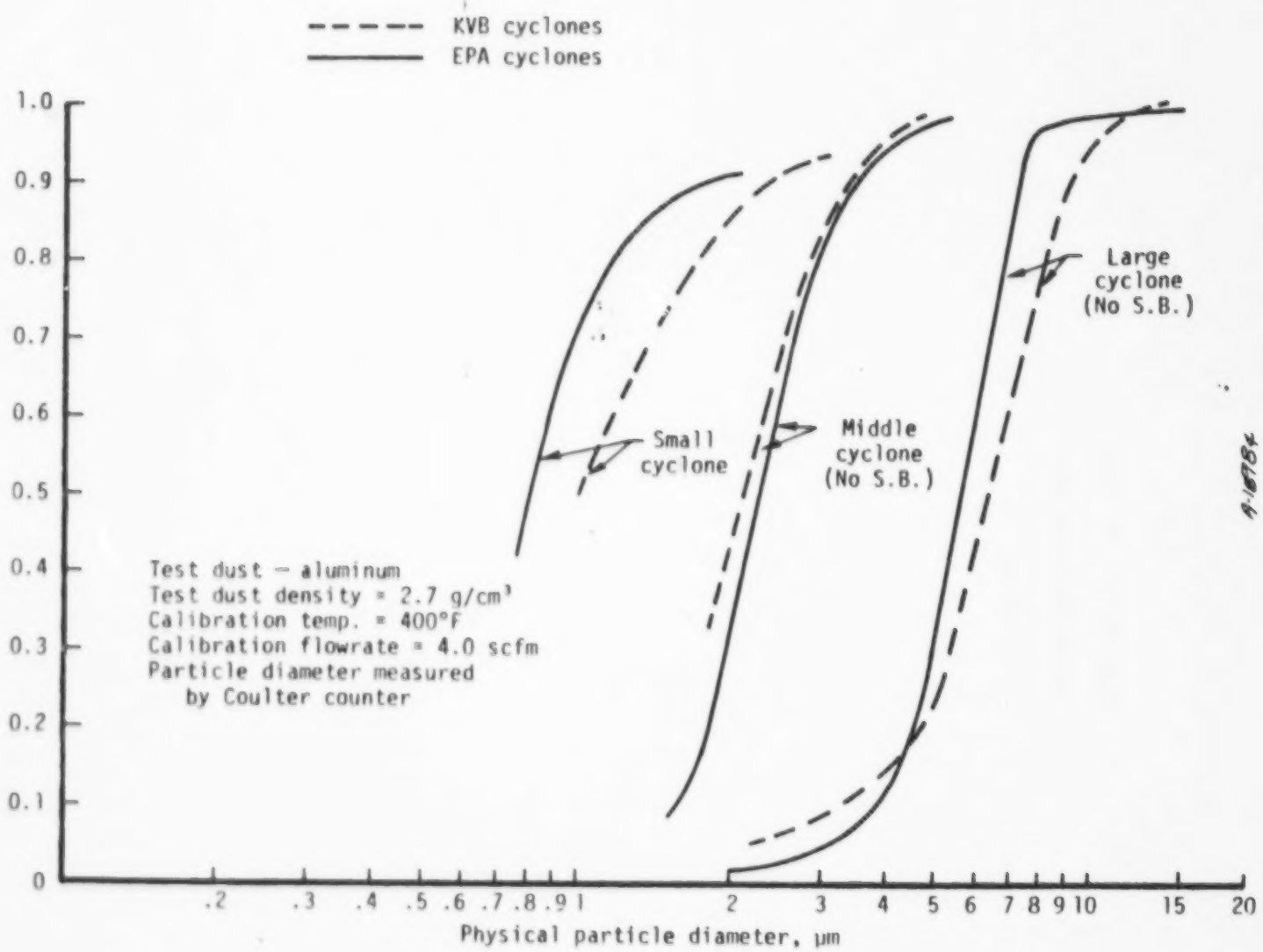


Figure 7-15. SASS cyclone calibration data.

PLANT OPERATIONS DATA

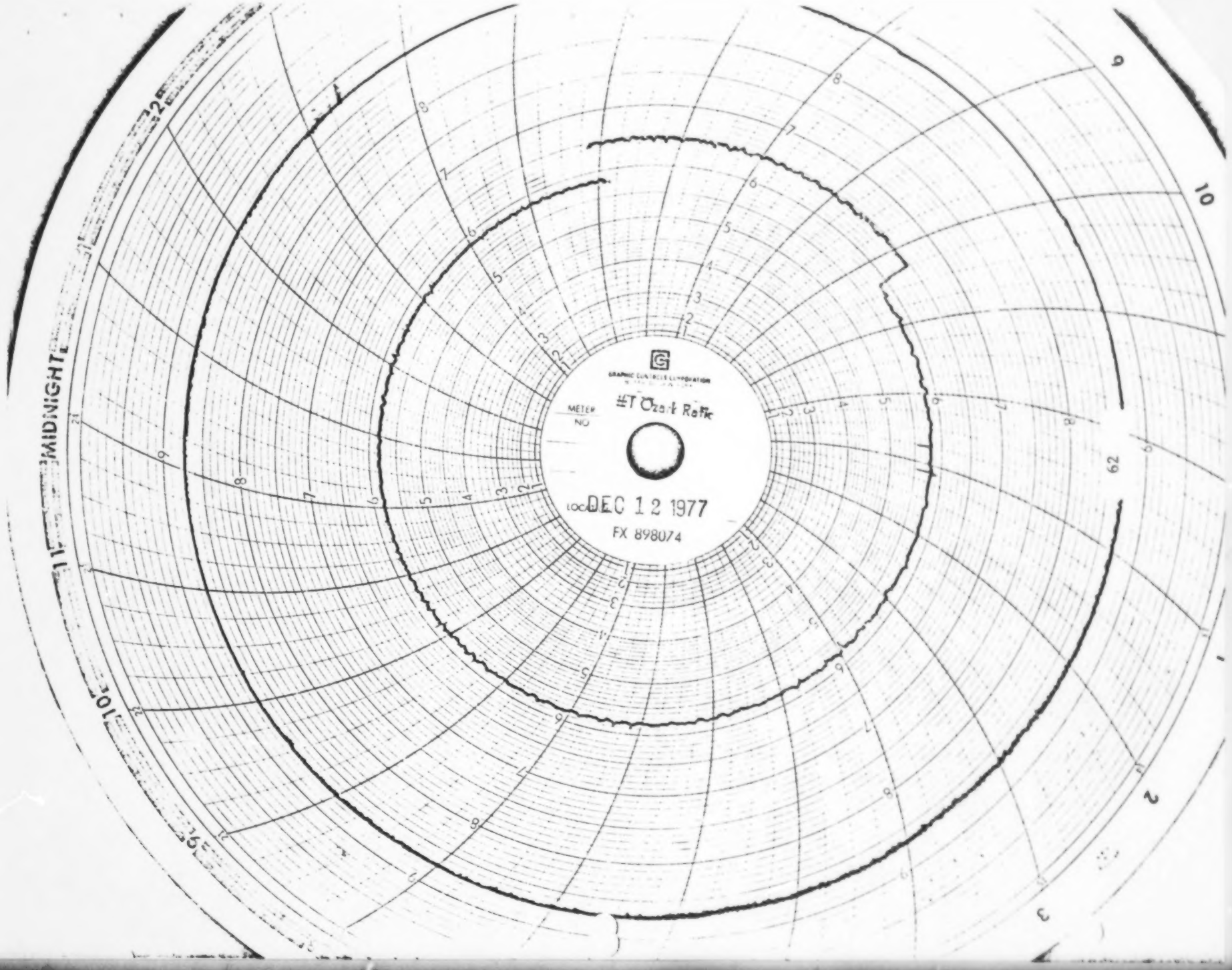
ORFICE PRESSURE DROP

STACK NO. 1

ORFICE PRESSURE DROP

STACK NO. 1

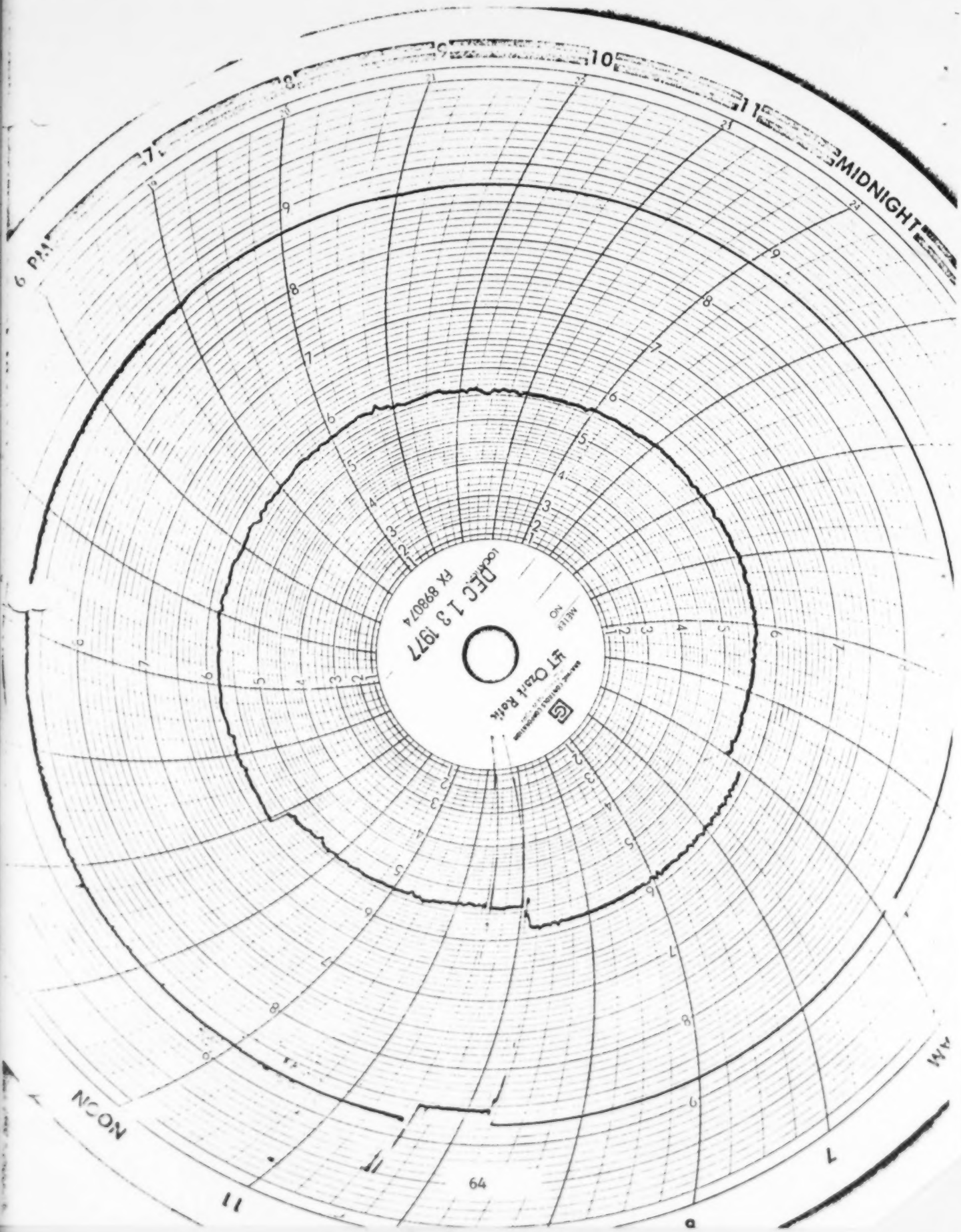
12/12/77


MIDNIGHT
METER NO
#T Ozark Refk
LOCATION
DEC 12 1977
FX 898074

ORFICE PRESSURE DROP

STACK NO. 1

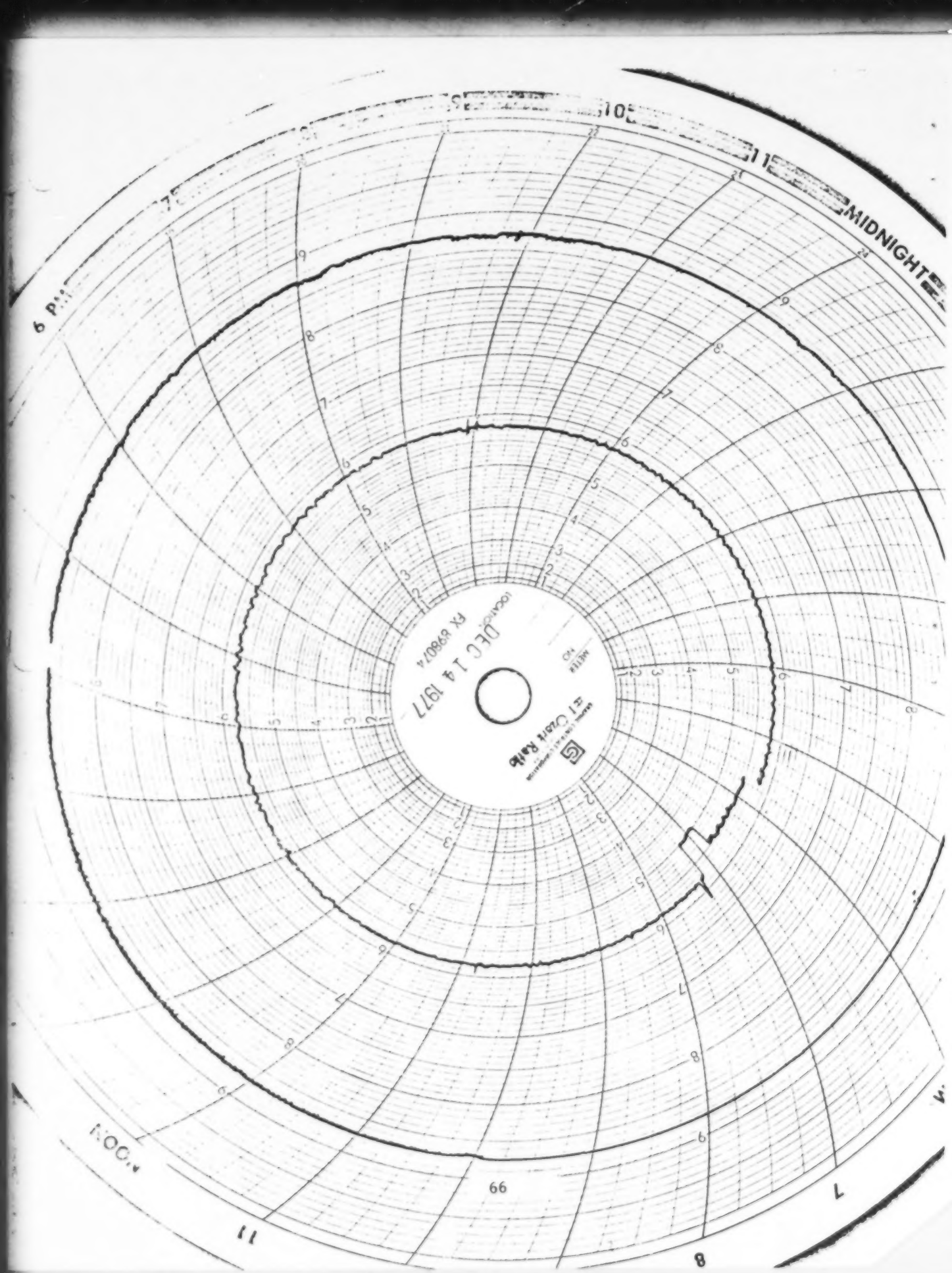
12/13/77


6 PM
MIDNIGHT
NOON
AM
DEC 13 1977
FX 898074
JT Ozark Rolk
64

ORFICE PRESSURE DROP

STACK NO. 1

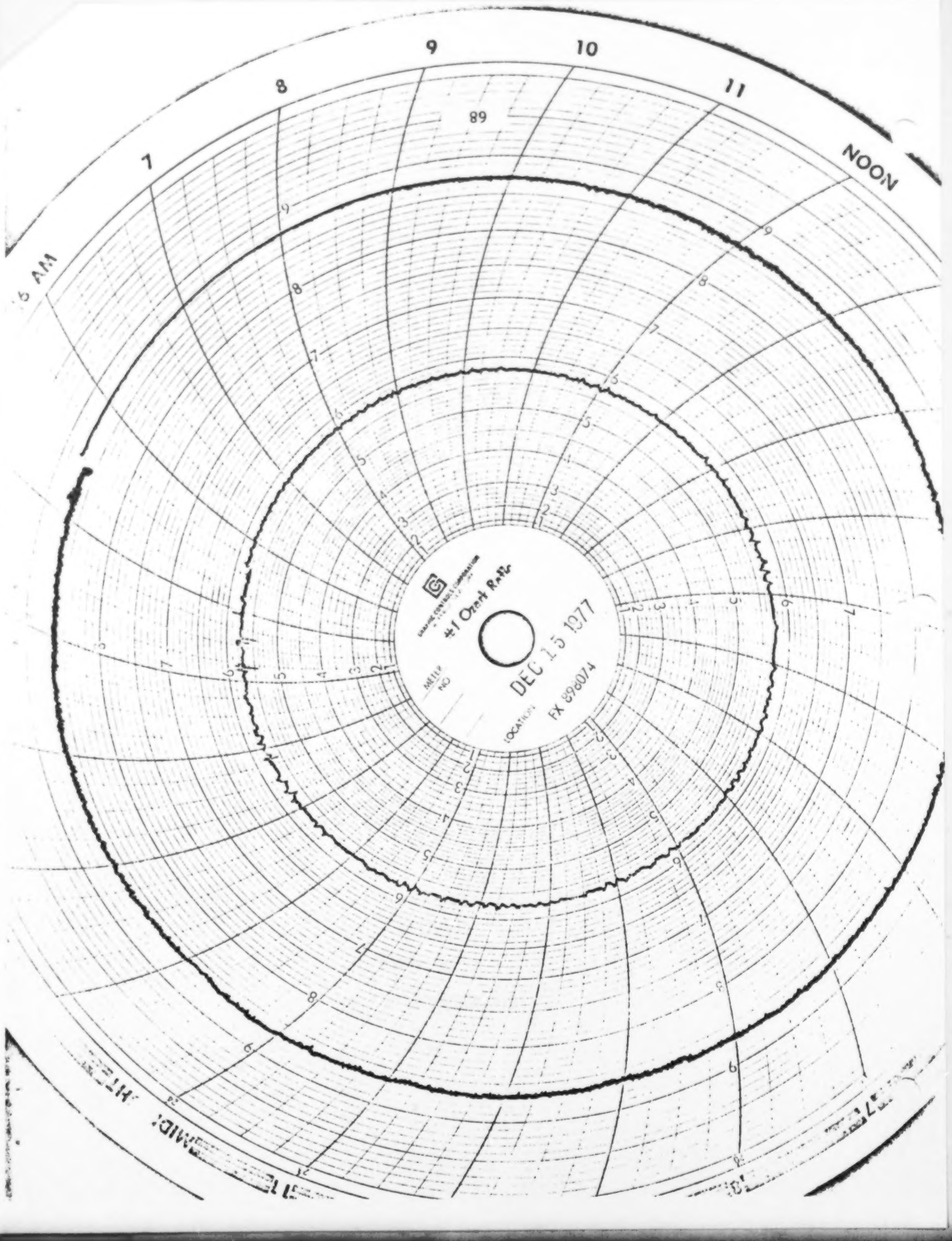
12/14/77


MIDNIGHT
DEC 14 1977
NOON

ORFICE PRESSURE DROP

STACK NO. 1

12/15/77


7
8
9
10
11
NOON
89
DEC 15 1977
FX 896074

SAMPLE CALCULATIONS

Particulate Isokinetic Sampling

1. DN = Nozzle Diameter, in.
2. P_B = Barometric Pressure, in. Hg
3. TT = Net Sampling Time, min
4. VM = VM final - VM initial = Sample Gas Volume, ft^3
5. TM = Average Dry Gas Temperature at Meter, ^oF

$$TM = \frac{\text{Avg TMI} + \text{Avg TMO}}{2}$$

6. PM = Average Orifice Pressure Drop, in. H_2O
7. Volume of dry gas sampled at standard conditions,[a] dscf

$$VMSTD = \frac{17.647 \times VM \left(P_B + \frac{PM}{13.6}\right)}{(TM + 460)}$$

8. VM = Total Water Collected - gm H_2O Silica Gel + ml Imp. H_2O
9. Volume of water vapor at standard conditions,[b] scf

$$VW \text{ gas} - 0.04715 \times VW = \text{scf}$$

10. Percent moisture in stack gas

$$\%\ M = \frac{100 \times VW \text{ gas}}{VMSTD + VW \text{ gas}}$$

11. Mole fraction of dry gas

$$MD = \frac{100 - \%M}{100}$$

12. Molecular weight of dry stack gas

$$MWD = (\%CO_2 \times \frac{44}{100}) \times (\%O_2 \times \frac{32}{100}) + \left[(\%CO + \%N) \times \frac{28}{100}\right]$$

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[a] Dry standard cubic feet at 68^oF and 29.92 in. Hg.

[b] Standard conditions at 68^oF and 29.92 in. Hg.

13. Molecular weight of wet stack gas

$$MW = MWD \times MD + 18\ (1 - MD)$$

14. AS = Stack Area, sq in. $= \left(\frac{\text{Stack diameter}}{2}\right)^2 \pi$

15. PS = Stack Pressure, in. Hg

$$PS = P_B \pm \text{Avg PST}$$

$$\text{NOTE: PST in. Hg} = \frac{\text{PST in. } H_2O}{13.6}$$

16. TS = Average Stack Temperature, ^{o}F

17. $SDE_{AV} = (\sqrt{\Delta p})_{AV} \times \sqrt{(TS + 460)_{AV}}$

18. Stack gas velocity at stack conditions, afpm

$$Vs = 5130 \times Cp \times \text{Avg (SDE)} \times \left[\frac{1}{PS \times MW}\right]^{\frac{1}{2}} = \text{afpm}$$

Where Cp = pitot tube coefficient

19. Stack gas volumetric flow rate at standard conditions,[a] dscfm

$$QS = \frac{0.1225 \times Vs \times AS \times MD \times PS}{(TS + 460)} = \text{dscfm}$$

20. Stack gas volumetric flow rate at stack conditions, acfm[d]

$$Q_a = \frac{0.05667 \times QS\ (TS + 460)}{PS \times MD} = \text{acfm}$$

21. Percent isokinetic

$$\% I = \frac{1039 \times (TS + 460) \times VMSTD}{VS \times TT \times PS \times MD \times (DN)^2}$$

22. Particulate, gr/dscf

$$\text{gr/dscf} = 0.0154 \times \frac{mg}{VMSTD}$$

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[a] Dry standard cubic feet per minute at 68°F and 29.92 in. Hg.

[b] Actual cubic feet per minute.

SASS TRAIN CALCULATIONS

1. ΔH_i setting for SASS train

   a. At $T_{oven} = 400^{o}F$, Q cyclones = 6.49 acfm

$$\Delta H_i = 1.888 \times 10^{-3} \left(\frac{Tm + 460}{Pm} \right) \left(\frac{1 - \frac{\%m}{100}}{.120} \right)^2$$

   b. At $T_{oven} = 200^{o}F$, Q cyclones = 6.49 acfm

$$\Delta H_i = 3.188 \times 10^{-3} \left(\frac{Tm + 460}{Pm} \right) \left(\frac{1 - \frac{\%m}{100}}{.120} \right)^2$$

2. Nozzle selection for SASS

$$D_N = 0.831 \sqrt{\frac{TS + 460}{VS \times PS}}$$

3. Cyclone Cut Diameters

$$D_{P2} = D_{P1} \sqrt{\frac{\mu_2 \, Q_1}{\mu_1 \, Q_2}}$$

Where D_{P1} = cut diameter at

$T_{oven} = 400^{o}$, Q_1 cyclone = 6.41 acfm

and $\mu_1 = 0.026$ centipoise

D_{P1} Small = 1.5 μm

D_{P1} Medium = 3.6 μm

D_{P1} Large = 10.2 μm

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